Chemical Engineering Progress

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LETTERS TO THE EDITOR

Sulfur Scrubbing

Gamson and Elkins' paper on Sulfur from Hydrogen Sulfide ("C.E.P." April, 1953) . . . and their development of theory and method of calculating equilibrium conversion is a distinct contribution. . . . They might have referred to the formation of monatomic sulfur at "free-flame" temperatures for an explanation of the higher-than-calculated equilibrium conversions obtained at high temperatures, particularly at the "free-flame" temperatures. The higher-than-calculated conversions obtainable in the low-temperature range is due, we believe, to published thermodynamic data . . . which give conversions lower than those obtained in the laboratory and plant.

Their references to basic developments in sulfur recovery have been unfair in failing to give credit to the development of the sulfur-scrubbing tower and circulation system for cooling and condensing the sulfur vapor and removing the mist formed in the condensation step. This was a development of the Southern Acid & Sulphur Company, who built and operated the first two commercial sulfur-recovery plants on the North American continent, using their own modifications of the Claus process. The two persons most responsible for this development were S. L. Nevins and J. S. Gilliam, both of Southern Acid & Sulphur Company, who developed the sulfur-circulation feature which made these pioneer operations a success. (Ref. U.S.P. 2,403,451 and U.S.P. 2,495,095) To-day there are hardly any commercially operating sulfur-recovery plants employing modifications of the "hot" Claus process which do not have a circulating sulfur-scrubber operation in their process. . . .

S. SCHWARTZ,
Consulting Engineer,
New York, N. Y.

LETTERS TO THE EDITOR

More on Registration

As of Feb. 1, 1953, there were 2,936 Professional Engineer registrants in Louisiana. This is exclusive of 132 registered as Land Surveyors and 317 in a separate "grandfather clause" classification of "Civil Engineers as Public Employees."

Of the 2,936 licensed in the various branches of professional engineering, 269, or 9.16%, are licensed to practice chemical engineering. Of these, 237 are registered in chemical engineering alone, and the rest are registered as follows: mechanical, 22; petroleum, 7; and one (1) each in metallurgical, civil, and mechanical-electrical.

In our New Orleans Section, A.I.Ch.E., 35 of 70 present members, or 50%, are registered. Of 48 members working or living in the city of New Orleans, 29, or 60.4%, are registered. The licensed members break down as follows: in chemical engineering alone, 29; in mechanical alone, 1; in chemical and mechanical, 4; in chemical and metallurgical, 1.

Compared with 1952, the above figures represent increases in total P.E.'s registered in Louisiana, the number and percentage in chemical engineering, the registered number and percentage registered in the N.O. Section.

Of the seven-man Louisiana Registration Board, two are chemical engineers and members of the Institute—Drs. Jesse Coates and Frederick W. zurBurg.

Registration news has appeared in news of various Local Sections in the past, as well as in separate news items in "C.E.P." Why not a section called "Engineering Registration News" or some equivalent title?

H. J. MOLAISON
Member, Professional Legislation Committee
New Orleans, La.

A New Name

I read with confused interest J. B. Austin's "petrocephalic counterfeiters" ["C.E.P." Noted & Quoted, July, page 10], suggesting the improper word "petrochemistry" for our new petroleum-derived chemical industry. Since Mr. Blaine said the Greeks had a word for it but didn't mention what it was, I wish to propose in the meantime or remind our members of a very appropriate word. This was recommended a year or so ago by the dean of petroleum technologists, Dr. Gustav Egloff. The term is "naphthochemistry," which should suit the most discriminating among us; the Greeks had nothing to do with it,—the root "naft," "neft," or "nift," meaning a burning oil is of Persian or Arabian origin.

Is everybody happy?

E. J. BARTH
Petroleum Technologist
Pittsburgh, Pa.

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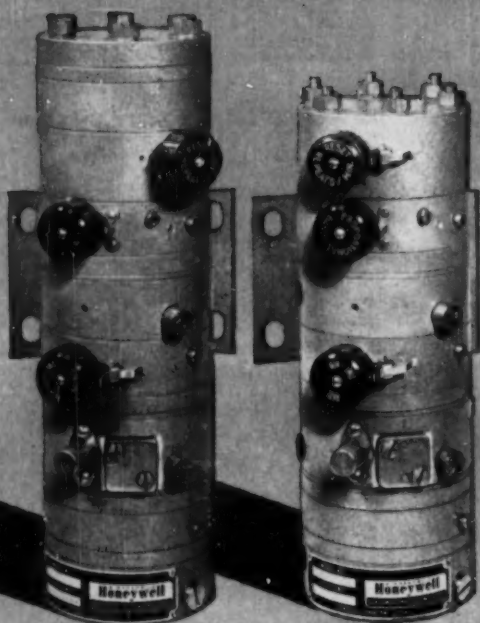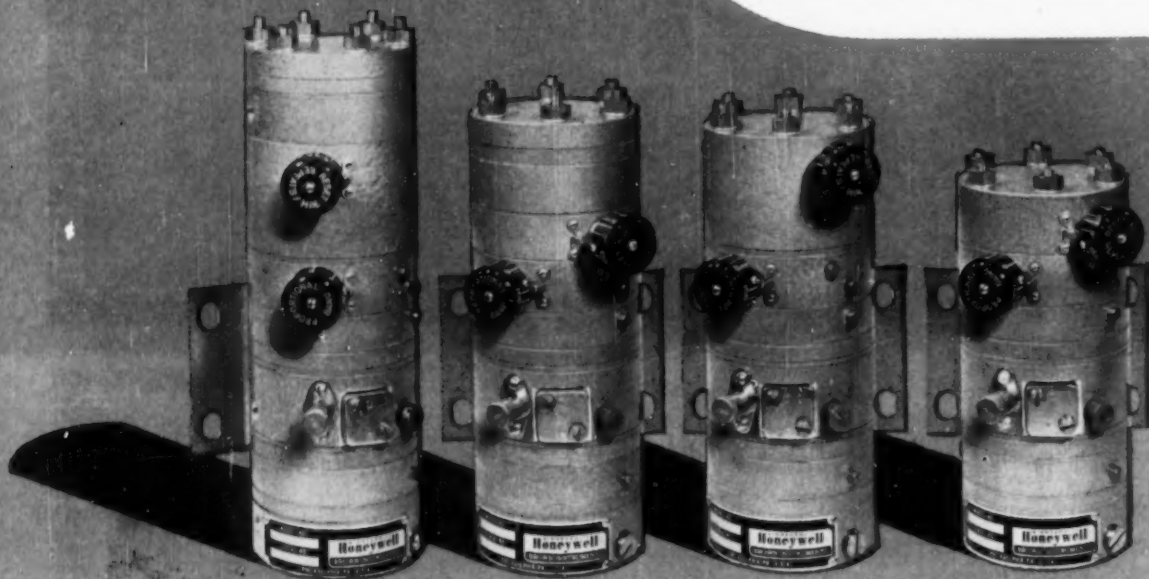
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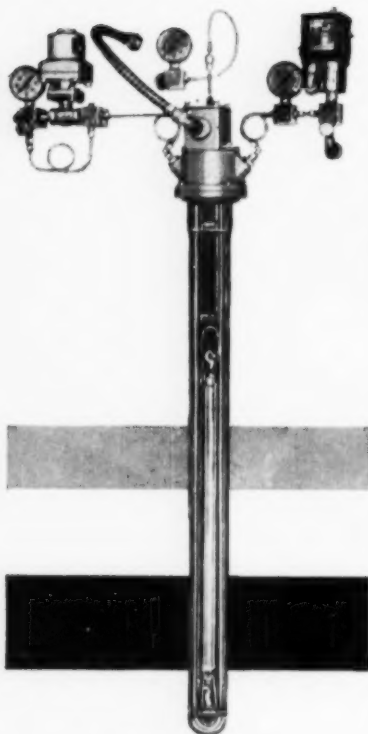
NOTED AND QUOTED

What Makes the Scientific Mind Scientific?

MAN'S accelerating progress in unlocking the doors to nature's mysteries has put a premium on scientific brains. As a discoverer of new horizons the scientist plays a crucial role in today's world. He and his works are in great demand.

Because so much depends on scientists and because the true scientist needs certain favorable conditions if he is to do his best work, it is important to examine the scientific mind and to know what are its characteristics. What qualities mark the scientific mind? How and to what extent does it differ from the minds of lawyers, writers, business men and others? How does the scientist do his job? . . .

Curiosity and Devotion

The most distinctive thing about a first-class scientist is that he is completely wrapped up in science: he eats it and sleeps it. The chances are that he works nights and Sundays and goes on vacation only at his family's insistence. . . . The scientist, more than most, finds in his work the mental and emotional rewards which should and can produce the greatest results and the greatest personal satisfactions.

These satisfactions stem from two qualities of the scientific mind. One is curiosity. The other is an ability to work out solutions objectively and gain new knowledge independently and in the scientist's own way. Scientists, above all, are people who have retained a childlike curiosity, which many of us lose as we grow up. They have wanted very much to find out about things and have learned that within certain limits they can, little by little, discover what they want to know. Moreover, they have learned that they can find out things *for themselves*—not by asking someone else or by looking it up in a book (granting that it could be learned that way, which is often impossible) but instead, thinking for themselves. . . . The need to become master of himself and his immediate environment is part of man's biological make-up. The scientist is one who has been fortunate enough to find a vocation which meets this need, for the gaining of knowledge is in itself a form of mastery, and deeply satisfying. Thus, the success of the scientist's work depends to an exceptional degree on his curiosity, love of the job and personal initiative. . . .

Freedom to Work

This assessment of the characteristics of the true scientist now brings us to the question: Under what circumstances does the scientific mind do its best work? It is evident that one of the most important elements in scientific work is the special freedom it gives its devotees to use their excellent minds to master knowledge in their own fashion. . . . But if science offers this special freedom it also requires special freedoms from and for the scientist.

No one can go far in science if he is not free to see things as they are. The man who is burdened by political or any other ideologies that restrict his thinking—whether they are accepted by him or imposed on him by society—is not free to examine the world to see if it is really in accordance with his beliefs. Men have not always and everywhere been free to become scientists. Creative productivity, not only in science but also in art and other fields, has varied greatly from culture to culture and from age to age in the same culture. From the periods and circumstances under which productivity has been high, it seems clear that an essential factor in any culture where creative activity flourishes is a high degree of independence and individual freedom. . . .

Given the qualities of the scientific mind—the driving curiosity, the personal approach to an enjoyment of the work, the high intelligence, the necessity for freedom—if the best creative faculties are to be used—it follows that management policies should be such as to cause a minimum of frustration to the scientific researcher. Enlightened management policies that give the scientist as much choice as possible in selecting his problem, that will make it possible for him to work in his own way and at his own time, and that are sensible in regard to security measures, would be a help.

Scientists do not regiment easily; they are stubborn and independent, and they have little patience with restrictions that seem to them unnecessary if not irrational. It is up to management and workers in other fields to realize these things about the scientific mind and make provision for them. In that way the nation can secure the best from its scientific brains.

Anne Roe
New York Times Magazine

Opinion *and* comment

DO WE TALK TO OURSELVES TOO LITTLE?

IN THE March issue of Chemical Engineering Progress D. H. Killefer's paper entitled "We Talk to Ourselves—Too Much" presented the public relations problem as it concerns the chemical industry and the scientist. Although I have no quarrel with Killefer's conclusions, I should like to approach this *talking* problem from another viewpoint—communications in an organization in which chemical engineers and chemists play a major role.

At a recent meeting of the New York Section of the A.I.Ch.E., T. E. Wannamaker presented an interesting discussion on "The Small Chemical Business." One of the major points in his talk was the noticeable difference in the attitude of chemical engineers and chemists toward their work in a small company compared to a large company.

In Dr. Wannamaker's own organization the technical people in the laboratory consider each project a personal challenge and work effectively to complete the job in the minimum of time. On the other hand, Dr. Wannamaker pointed out that in many large research laboratories in the country, he found the engineers and chemists more interested in hovering around the desk of a good-looking stenographer or in making preparations for departure well in advance to be ready to leave the laboratory bench promptly at, or even slightly before, quitting time. He ascribed this difference in attitude to the difference in size of the organizations. This is probably a logical conclusion, but I doubt that size alone is the whole answer.

Analysis of this problem, reveals the basic cause to be that communication is much more difficult in a large organization. The engineer or chemist is often given only a small part of an over-all problem to work on and the background of the problem is not explained. He is not kept informed of the progress made on other segments of the problem. The tendency is, all too often, to pass on to the individual only the minimum of information required for his part of the work, rather than to tell him why the work is necessary and what the benefits of a job well done will be.

While I do not propose any cure-all for this situation, I believe that the following ideas might be of some help:

1. When an assignment is *given*, questions should be asked about the background, why the job is necessary, and why it has to be done by a certain date. By showing an interest in these facts, encouragement will be given those issuing the assignments to impart all the facts as a matter of course.

2. When *giving* an assignment, the following questions should be asked:
   (a) Have I given all the information necessary for the execution of a good job without wasted effort on nonessentials?
   (b) Have I explained the purpose of this assignment and the results which might be expected?
   (c) Have I stated the time available for the completion of this task and the reasons for this limitation?
   (d) If others are woring on other portions of the job, have I explained what their assignments are, so that there is no overlapping and so that the individual to whom this assignment is given can check, when necessary, with others doing other parts of the job?

3. Conferences should be set up periodically for exchange of information on the progress of different projects or portions of projects. Such meetings or panel discussions will not only add to each individual's interest in the project, but will actually eliminate overlapping work.

4. Asking questions should be encouraged and all questions should be answered as completely and informatively as possible.

Let's talk to one another more!

L. P. SCOVILLE

Operating Problems in Ammonia Synthesis

A. V. Slack, H. Y. Allgood, and Harold E. Maune

Tennessee Valley Authority, Wilson Dam, Alabama

Every so often out of the blue, comes a manuscript which completely describes some phase of chemical engineering . . . an article important from a sheer operating standpoint . . . the editor wonders why chemical engineers, so vastly concerned with operating plants, don't write more articles such as the one we print here on operating problems . . . this is about the synthetic ammonia field now enormously important and producing about two million tons a year . . . while this paper describes a particular process, its significance is in the fact that it describes an operating procedure.

A. V. Slack, now chief of exploratory research in the Tennessee Valley Authority's division of chemical development at Wilson Dam, Ala., joined T.V.A. in 1941 after receiving a B.S. degree from East Tennessee State College and an M.S. from the University of Tennessee. As a laboratory supervisor he took part in the initial operation of the T.V.A. ammonia plant in 1942-43. In subsequent assignments as staff chemical engineer, he has worked on various design and procurement problems in connection with ammonia and nitric acid production. He has been active in the organization of industry-sponsored schools for teachers sponsored by the A.I.Ch.E. Chemical Engineering Education Projects Committee.

H. Y. Allgood is a chemical engineer in charge of ammonia plant studies in the Tennessee Valley Authority's Plant Chemical Control Branch. He joined the T.V.A. in 1936. His work has been in research, ammonia plant control and operation, catalyst preparation and evaluation, and process studies. He received his A.B. and M.S. degrees from Emory University.

Harold E. Maune, now plant chemical engineer of the ammonia branch of the Tennessee Valley Authority's Division of Chemical Operations at Wilson Dam, Ala., joined T.V.A. in 1942 as a chemical engineer. His work has been concerned with ammonia plant operations. He received his B.S. degree from Missouri School of Mines and Metallurgy in 1942.

Of the many applications of high pressure in chemical processing, ammonia synthesis is one of the oldest and best developed. From the first full-scale production in 1913, direct catalytic synthesis from hydrogen and nitrogen has been developed until it is now a fairly well-standardized operation. Growth in the industry has been especially pronounced in the past decade; production capacity in this country has increased from about 570,000 tons/year in 1941 to approximately 1,170,000 tons in 1951. The level predicted for 1955, in view of current expansion plans, is 2,650,000 tons. This should qualify ammonia synthesis as one of the fastest growing heavy chemical industries.

Process descriptions and design features of ammonia plants have been the subject of a number of papers in recent years. Most of these are fairly general, with principal emphasis upon existing design practice. As an extension of such information, a case history of an operating plant, with emphasis on the specific problems encountered in a high-pressure process, may be of value and may be helpful in relating design considerations to efficient plant operation and low-maintenance cost.

History of T.V.A. Ammonia Plant

This paper tells the story of the Tennessee Valley Authority plant operation at Muscle Shoals, Ala. Built in 1941-42, this plant has been described previously (1, 3, 5). The following description is limited to a summary of the process and details of performance not covered in previous papers.

A modified Haber-Bosch process is used in the plant. Synthesis gas was produced originally from coke in semi-water gas generators, but in 1951 the plant was converted to use of natural gas, a more economical raw material than coke. In either case the synthesis gas must be treated to remove carbon oxides, which is done by successive scrubbing under pressure with water and an ammoniacal solution of copper salts. The purified gas then enters the synthesis units at 200 to 350 atm. pressure.

A flow sheet for the synthesis circuit is shown in Figure 1. The purified
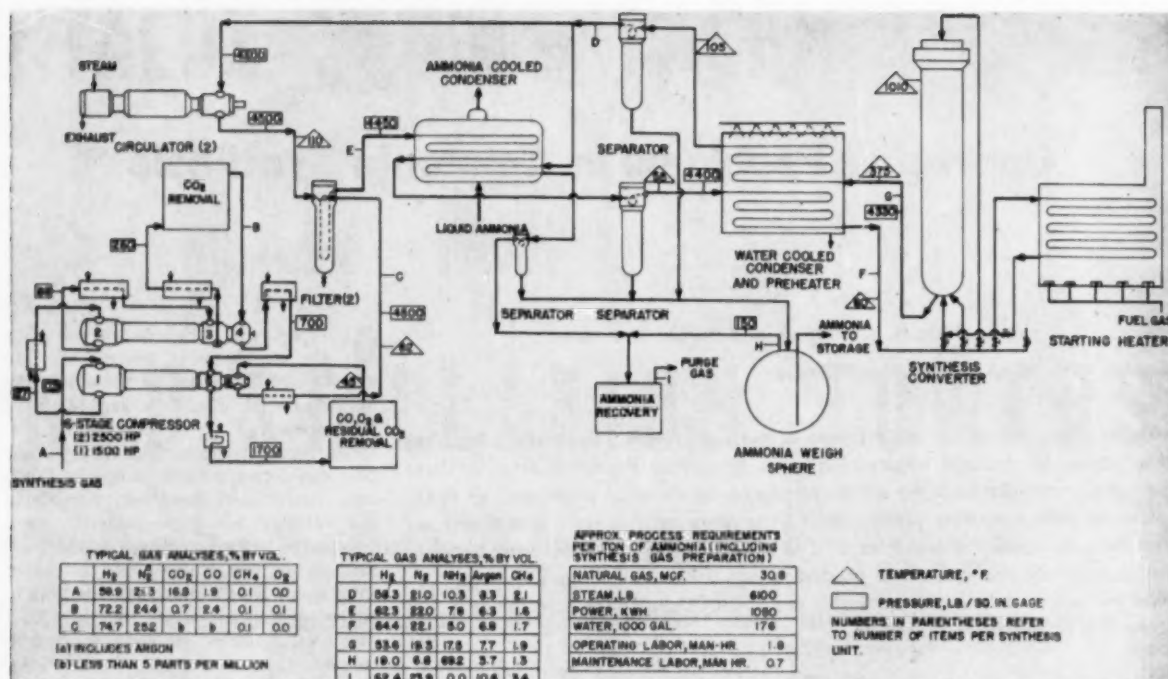
Fig. 1. T.V.A. ammonia synthesis unit.

TYPICAL GAS ANALYSES, % BY VOL.

	H_2	N_2	CO_2	CO	CH_4	O_2
A	58.9	21.3	16.8	1.9	0.1	0.0
B	72.2	24.6	0.7	2.4	0.1	0.1
C	74.7	25.2	b	b	0.1	0.0

(a) INCLUDES ARGON
(b) LESS THAN 5 PARTS PER MILLION

TYPICAL GAS ANALYSES, % BY VOL.

	H_2	N_2	NH_3	Argon	CH_4
D	58.3	21.0	10.5	8.3	2.1
E	62.5	22.0	7.6	6.3	1.6
F	64.4	22.1	5.0	6.8	1.7
G	53.6	19.5	17.5	7.7	1.9
H	19.0	6.8	69.2	3.7	1.3
I	62.4	23.9	0.0	10.6	3.4

APPROX. PROCESS REQUIREMENTS PER TON OF AMMONIA (INCLUDING SYNTHESIS GAS PREPARATION)

NATURAL GAS, MCF.	30.8
STEAM, LB.	6100
POWER, KWH.	1080
WATER, 1000 GAL.	176
OPERATING LABOR, MAN-HR.	1.8
MAINTENANCE LABOR, MAN HR.	0.7

△ TEMPERATURE, °F.

▢ PRESSURE, LB./SQ. IN. GAGE

NUMBERS IN PARENTHESES REFER TO NUMBER OF ITEMS PER SYNTHESIS UNIT.

Fig. 2. Ammonia synthesis converter.

make-up gas mixes with recirculating synthesis circuit gas in the filter, the purpose of which is to remove oil and other nongaseous impurities. The mixed gas then passes through an ammonia-cooled condenser, an ammonia separator, and a preheater on the way to the converter. In the converter (Fig. 2), the gas flows through a heat exchanger in the lower section, up through catalyst cooling tubes in the upper section, down through the catalyst, and finally through the exchanger to the converter outlet. Part of the ammonia produced is removed in the water-cooled condenser before the gas passes through the circulator compressors to complete the circuit at the point of junction with the make-up gas.

The synthesis section is made up of two identical trains, with two filters and two circulators for each. Train 1 was started up in August, 1942, and train 2 in September of the same year. Both trains were operated at a production rate of about 90 tons/day each for several years without opening the converters. In late 1949 train 2 began failing rapidly with evidence both of loss of catalyst activity and of need for repair of converter internal parts. The train was shut down in February, 1950, and the converter thoroughly renovated. The catalyst basket assembly was replaced with a new one, the heat exchanger was retubed, and new catalyst was charged to the unit. When the train was returned to service, performance was as good as

at the beginning of the first catalyst run and has continued so since.

Train 1 continued to operate satisfactorily after the failure of train 2. In early 1951, however, the plant production was reduced by about half during the installation of equipment for use of natural gas. Since train 1 had clearly lost some of its original efficiency, although still capable of its rated capacity, it was decided to rehabilitate the converter during this period while train 2 alone could carry the load, thus avoiding interruption in production later when the converter eventually failed. A new basket and exchanger were installed and the unit returned to production in July, 1951. Operation since then has been satisfactory.

In planning the conversion to natural gas it was found by calculations based on plant data that with minor changes all the plant units except the compressors could be operated at levels considerably higher than the nominal design rate. Two new compressors were installed and the synthesis trains have since produced at rates of 120 to 140 tons/day each, with an average of about 125 tons. It is not expected, however, that the long catalyst runs of 7½ years on train 2 and 8½ years on train 1 will be experienced at the higher rate.

Compression and Circulation of Gas

As might be expected, some of the most troublesome operating problems in the high-pressure system arise in connection with the moving parts of equipment such as compressors, circulators, and valves. The original compressor installation consisted of four six-stage Cooper-Bessemer reciprocating compressors of 100 rev./min., tandem-cylinder type, each of which was driven by a 2500 hp. synchronous electric motor mounted on a common frame. The two new Clark compressors installed in 1951 are of the horizontally opposed type and operate at 300 rev./min.

Breakage of valve springs and plates is the difficulty most frequently encountered, caused principally by entrained water in the gas. The gas leaving the cooling towers in the gas production section is saturated with water, and in each compression stage followed by cooling in the intercooler causes water to condense in the gas stream. When the valves close on the resulting droplets, stresses are produced which cause breakage. The intercoolers are so constructed that most of the water is drained directly from them. However, it has been found necessary to install separators of various types after all the intercoolers to minimize the amount of water passing on into the cylinders. In addition, a separator was installed on the main line to the compressors to remove scale, mist, and other contaminants.

These changes have improved compressor operation to the extent that valves in the lower pressure stages now last as long as four years.

In the higher stages, however, valve breakage remains the most frequent cause of compressor down time. The effect of this is minimized by reducing the time required to change a valve; by close cooperation between operating and maintenance personnel it is possible to make the change in as little as five minutes.

Gas leakage through piston-rod packing has presented no serious problems. Carbon-Bakelite packing is used for operation up to 250 lb./sq.in. and bronze packing for higher pressures.

Next to valve trouble, corrosion by cooling water has been the most troublesome operating problem. For cooling the cylinder jackets, a system is used in which cooling water is circulated to the jackets, to a surge tank, through a cooler cooled by process water, and back to the jackets. Both ordinary filtered water and Zeolite-treated boiler water were found to be excessively corrosive in this system; chromate-treated filtered water was then tried and found satisfactory. The chromate concentration is maintained at 700-1000 p.p.m. and the pH at 7.5 to 8.5.

Corrosion has been a problem also in the intercoolers, which are of double-tube type with the gas flowing through the annular space. In the original installation finned inner tubes of mild steel were used in the first three compression stages. These were corroded by condensate in the gas stream, with the result that the fins would occasionally break off and pass on into the compressor cylinders. The resulting valve breakage made it necessary to remove the fins. In addition, the mild steel tubes were replaced with stainless steel ones to reduce corrosion.

An initial problem was the handling of the piston blow-by gas on single-acting pistons in the higher stages. This gas collects behind the piston and must be vented from the cylinder. Originally, it was piped back to the compressor inlet. However, since the first-stage-inlet capacity controlled the over-all compressor capacity, addition of the blow-by gas at the first stage was undesirable. It was possible to add the gas at the inlet to the second stage, thereby effectively increasing the compressor capacity because stages two to six inclusive were capable of handling the additional load without overloading the compressor or affecting the amount of gas entering the first stage. As a result ammonia production was increased appreciably.

In addition to the above unscheduled causes for compressor down time, regular shutdowns are made for maintenance inspection and cleaning of intercoolers and motors. Every effort is made to keep compressor down time at a minimum, since the plant operates without a spare compressor. Table 1 indicates the amount of operating time the ammonia plant lost which can be attributed to compressor repairs and maintenance. Some additional compressor maintenance is done in periods when the ammonia plant is down for reasons other than compressor trouble, but there has been relatively little opportunity for this, since other units of the plant are seldom out of operation on a planned schedule.

Fig. 3. Filter assembly.

Table 1.—Compressor Operation

Year	Down Time, %[1]
1943	1.1
1944	0.6
1945	0.6
1946	0.7
1947	0.8
1948	1.1
1949	1.0
1950	1.9[2]
1951	0.9

[1] Total down time attributable to the compressor section expressed as per cent of total time the compressors would otherwise have operated.
[2] Motor failure occurred during this year.

The synthesis circulating compressors have given good service. During initial operation excessive leakage occurred at the packing because of piston-rod wear. The pistons were replaced with harder rods, forged of S.A.E. 4620 steel and hardened to 60 Rockwell C. The replacements, in conjunction with a good grade of bronze packing, have given good service.

Valves in the T.V.A. plant are mainly of the angle type. Most of the larger sizes are motor-driven, and these have given some trouble because of slippage of the motor setting, resulting in the motor's moving the stem beyond its open limit and thereby snapping the stem off. This necessitates constant checking of the motor setting.

The major valve troubles have been with the packing. Diamond-cut metallic packing was used originally, with 8-in.-deep stuffing boxes on the large valves. Leakage was excessive, and the valves were hard to turn. Teflon V-ring packing

GAS OUTLET

MAKE-UP GAS INLET

GAS INLET

FILTER ELEMENTS

RASCHIG RINGS

DRAIN
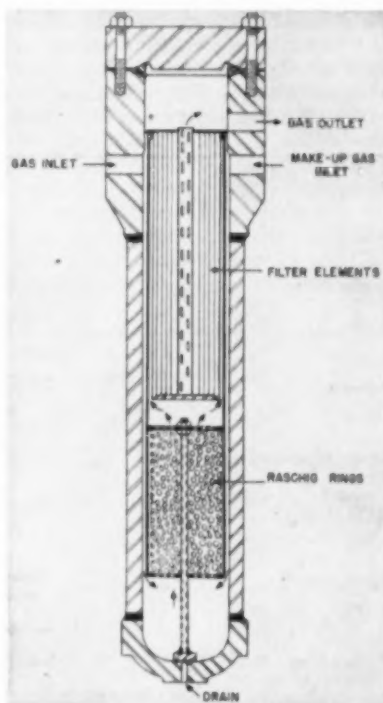
has been found to be superior, and it is now being installed in all valves. The valves are more easily moved, and leakage has been eliminated. A 3-in. depth of the Teflon packing is sufficient.

Valve seats and plugs are made of nitrided steel and are renewable. At points where the valve stays in one position most of the time, many of the original seats and plugs are still in service. The most severe service is in the filter outlet valves, which must be opened and closed every 48 hours; renewal of seats and plugs in these valves is necessary about every two years.

Filtration of Gas

One of the more troublesome operating problems is removal of solid and liquid impurities from the circulating gas in order to prevent contamination of the ammonia produced, plugging of lines, and injury to the catalyst. The impurities are mainly of three types: oil from the compressors and circulators, scale and dust from piping and equipment, and compounds formed by reaction of ammonia in the circulating gas with gaseous impurities in the make-up gas. Oil contamination is kept to a minimum by using a heavy oil and through careful control of the amount used. However, even under best operation a significant amount escapes into the gas stream and must be removed. The main source of compounds formed in the circuit is residual carbon dioxide in the make-up gas. This reacts with ammonia to form compounds which are especially troublesome because of their tendency to plug equipment. Ordinarily, the carbon dioxide content of the entering make-up gas is kept below 5 p.p.m., but it may go as high as 10 p.p.m. if the copper liquor absorption unit in the purification section is operating under heavy load. Even

at the lower concentration, however, sufficient carbonates are formed in the synthesis circuit to make their removal from the system desirable.

The filter unit used is shown in Figure 3. Make-up gas and recirculating gas enter near the top of the unit and flow downward through an annular space between a sleeve and the shell. The gas mixture leaves the annular space near the bottom of the filter and flows upward through a basket that contains mild steel Raschig rings. The gas continues upward into an annular space between the sleeve and the filter element, from which it flows inward through the element and into a slotted outlet pipe. The filter element consists of six concentric screen units sealed at the top and at the bottom so that gas must flow through all the units. Each of the five outer units consists of layers of stainless steel wire screen supported on ribs; the inner unit is the slotted gas outlet with one layer of screen. One or more of the outer units is covered with a filter cloth. When the plant was initially put into operation in 1942, spun-glass filter cloth was used. Since this became plugged rapidly with the oil-carbonate residue, and since it did not readily tear, resistance to gas flow increased until the filter element collapsed. Cotton waste material inserted between the concentric units was tried instead, but it failed also. The material finally adopted was common cotton domestic cloth (shrunk), which retains the residue satisfactorily and on sealing over usually rips before pressure becomes so great as to collapse the element.

Two filters are used in parallel on each train; one is in service while the other is idle for in-place cleaning. A filter is operated continuously for 48 hr. and is then closed off and cleaned by steaming. After a thirty-day service, a filter is opened and the element is removed, thoroughly cleaned, and repaired if necessary. Because of corrosion the Raschig

rings require replacing every one to two years.

The filters remove oil quite effectively, down to less than 2 p.p.m. in the ammonia produced. Removal of carbonate was also satisfactory at the original level of production, although enough carbonate passed through and collected in downstream piping to necessitate a shutdown about every nine months for steaming out coils in the ammonia-cooled condenser and the preheater. Such a shutdown involves a production delay of 4 to 5 hr.

The increase in production to 125 tons/(day)(train) has caused some overloading of the filters, increasing the frequency of filter element failure and of shutdowns to steam out condensers. Figure 4 shows a set of filter elements, the outer ones of which collapsed because of differential pressure through the filter. To alleviate this condition a centrifugal-type separator will be installed in the filter in the space now occupied by Raschig rings. Some relief is expected also from changes in the purification system designed to reduce the amount of carbon dioxide entering the synthesis circuit. Should these prove inadequate, it may be necessary to install auxiliary oil separators in the lines entering the filter.

Control of Process Variables

The main objective in control of process variables in high-pressure ammonia synthesis, as in any other process, is to produce the required amount of product with a judicious balance between low operating cost, low risk of equipment damage and process interruption, and long life of equipment. An important function of the operating department is to determine the optimum bal-

Table 2

	HIGH	LOW	OPERATIONAL CONTROL
Temperature	Loss of catalyst activity	Lower production rate[1]	Variation of space velocity
	Shorter life of converter parts		By-passing of converter heat exchanger
	Lower production rate[1]		
Pressure	Equipment limit	Lower production rate	Adjusts automatically to a particular combination of temperature and space velocity
Ammonia returned to converter	Lower production rate	Refrigeration equipment limit	Regulation of amount of refrigeration used
Inerts in circuit	Lower production rate	Greater loss of synthesis gas in purging	Bleeding off of synthesis gas, with ammonia recovery (some unavoidable loss by leaks)
Space velocity	Poorer thermal control of converter	Lower production rate	Variation of circulator speed
	Refrigeration equipment limit		
	Increased cost of gas circulation		

[1] For any given catalyst and space velocity, there is an optimum temperature, above or below which production falls off.

ance between these factors. In ammonia synthesis the problem is complicated because of the fact that a general decline in process efficiency must be accepted as the catalyst loses activity. This necessitates changes in the balance between the various process variables during the life of the catalyst.

The variables involved are temperature, pressure, space velocity (volumes of effluent gas at standard conditions per bulk volume of catalyst per hour), ammonia content of gas entering the converter, content of inert constituents in the synthesis circuit, and hydrogen-nitrogen ratio. Details of the effects of variables on reaction equilibrium are available elsewhere (1, 2, 7). In general, a high degree of conversion is favored by low temperature, high pressure, low inert gas content, and close approach to a hydrogen-nitrogen ratio of 3.0. In practice, however, it is necessary to depart somewhat from conditions favorable to equilibrium. Limiting considerations which affect both design and operation are as shown in Table 2:

Proper operating balance between variables is affected to a great extent, of course, by the amount of catalyst in the converter. From the operating standpoint, use of a generous amount of catalyst has a number of advantages. It allows a relatively wide variation in individual process variables, helping to offset any shortcomings which may exist or develop in other equipment in the synthesis circuit. Moreover, the large body of catalyst serves as a heat reservoir during temporary interruptions in production caused by trouble in other parts of the plant. In many cases this makes it unnecessary to use the starting heater to get back to full operation, which is a slow and delicate operation.

The principal advantage of the large catalyst body, however, is the longer period of operation before recharging becomes necessary. In a plant of the T.V.A. type at least ten days are required to shut down and cool the converter, replace catalyst, and start up again. And after getting back on stream, a day or so usually elapses before the process can be fully tuned and brought back to full efficiency. Moreover, thermal stresses, accompanying the cooling and reheating, affect the life of the equipment and should be avoided, if possible.

In the T.V.A. plant the relatively large amount of catalyst installed has made it possible to operate until the converters needed shutting down anyway for internal repairs, thus making one shutdown cover both needs. Each of the converters contains about 144 cu.ft. (about 12 tons) of catalyst. This is roughly about twice as much catalyst as is used in some of the other 350-atm.
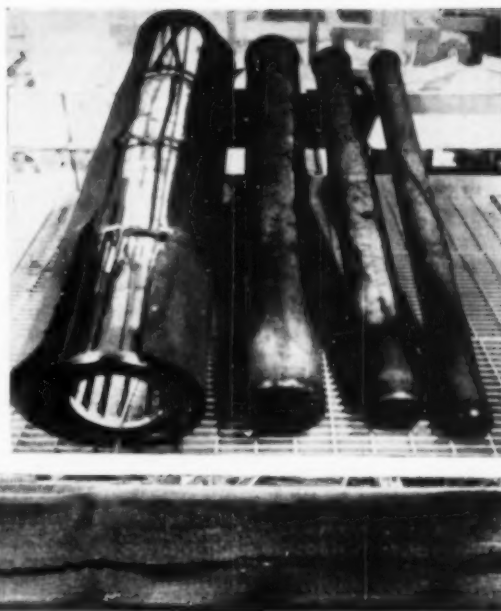
Fig. 4. Filter elements removed from recirculating gas filter; outer element (No. 6) in foreground; inner elements at rear (Nos. 5, 4, 3, 2, and 1); Nos. 5 and 4 are still assembled.

processes (3). However, since the cost of the T.V.A. converters represented only about 3% of the total ammonia plant cost, the larger size of the converters was not a significant cost item, and the intangible value of long-term, uninterrupted operation is considered to be well worth the small additional cost involved.

The range of operating variables involved in maintaining production for more than eight years with the same catalyst is indicated in Table 3. Data given are typical ones from operation of train 1 and were selected from periods of approximately equal production rate. Also shown, to illustrate the shift in variables required to get higher production, are data taken near the beginning of the second catalyst run on train 1. At the beginning of this run production was increased to about 125 tons/(train)

(day) as a result of the conversion to natural gas.

The most important single operating variable is the temperature in the catalyst bed. On this depends, to a great extent, both the life of the catalyst and the internal parts of the converter. Every effort is made to keep the temperature at as low a level as is practicable and also to prevent wide variations in temperature within the catalyst bed.

In the T.V.A. converters, catalyst temperature is measured by three thermocouples, the wells for which extend downward through the catalyst mass and are located in line on a diameter—one at the center and each of the other two about 10 in. from it. The thermocouples are adjustable vertically; the center one is normally kept at the point of maximum temperature, the control point for operation. The other two register inlet

Table 3.—Typical Converter Operating Data

SPACE VELOCITY	PRESSURE,[1] ATM.	MAXIMUM TEMP., ° F.	% INERTS IN CIRCULATING GAS	% NH₃ IN GAS TO CONVERTER	% NH₃ IN GAS FROM CONVERTER	NH₃ PRODUCED, TONS/DAY	COMMENTS
9050	245	997	11.1	4.0	17.9	93	Near beginning of first catalyst run on train 1; train started 8-4-42
12700	328	1051	7.9	4.0	13.8	91	Near end of first run; train shut down 2-5-51
13200	263	991	11.1	5.3	18.1	122	Near beginning of second run on train 1

[1] At converter outlet.
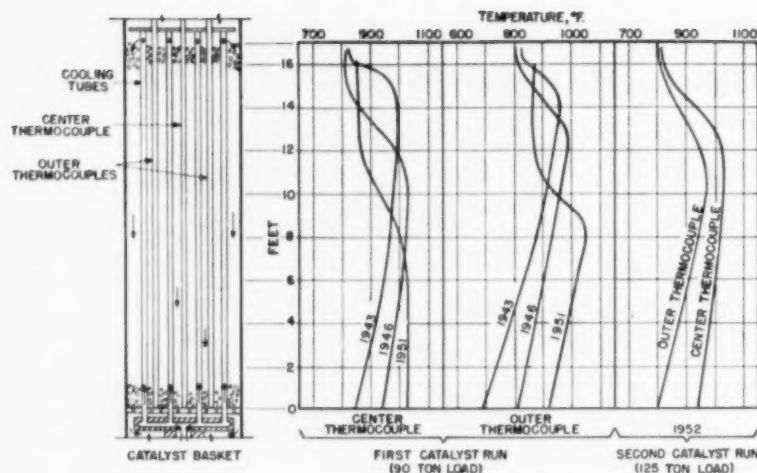
Fig. 5. Typical temperature distribution in catalyst.

and outlet temperatures. In addition, the vertical temperature gradient is determined periodically.

The range of temperatures encountered in the converter is illustrated in Figure 5. One of the sets of data shown was taken in January, 1943, shortly after the beginning of the first catalyst run on train 1, the second in 1946, and the third in January, 1951, shortly before the end of the run. With the active catalyst in 1943, the curves are characterized by fast heat release as soon as the incoming gas strikes the catalyst. The maximum temperature is reached in about the first 2 ft. of travel through the catalyst; thereafter, transfer to the cooling tubes becomes faster than heat release from the continuing reaction, so that a drop of 200° F. or more occurs down through the remaining 14 ft. It would be better, of course, to have cooling tubes designed so that transfer of heat would be faster at the hot point than at other points, thereby evening out the catalyst temperatures. This appears impracticable, however, since the hot point does not remain in the same place throughout a catalyst run. Figure 5 shows that the point of maximum temperature had dropped 3 or 4 ft. by 1946 and was down in the lower half of the converter by the end of the run. In view of this, the best design appears to be a simple straight tube with a core tube to give maximum transfer rate.

Although the hot point was at about 1000° F. at the beginning of operation, the bulk of the catalyst was below 900° F. and a considerable portion below 800° F. It would have been possible to operate at an even lower temperature, since both space velocity and pressure were at relatively low levels (see Table 3) and these could have been increased to balance the adverse effect of lower

temperature on space-time yield. However, the rate of conversion drops off fast as the temperature is lowered below the 800-900° F. range, with the result that the reaction becomes difficult to control. At the higher temperature levels an accidental off-balance in a process variable such as a high hydrogen-nitrogen ratio ordinarily would affect operation only by causing the hot point to move downward a little in the converter. At lower temperatures, however, the same deviation might "blow out" the reaction with the result that catalyst temperature and production would fall off, and general re-tuning of the unit would be required. For this reason the temperature is kept high enough to cushion the effect of minor changes in process conditions. The higher temperature also allows lower space velocity and pressure, with a consequent economy in compression and circulation.

As the catalyst gets older, an increase in average temperature is required, not only to give the temperature cushion, but also to aid in maintaining production after the other variables have been pushed to their limits. Figure 5 shows that the increase in maximum temperature was held to about 50° F. over the 8½-year run, although the average temperature of the catalyst increased considerably more as is indicated by the increase of about 200° F. at the outlet.

The problem of temperature control in the converter is complicated by the necessity of keeping the converter shell cool. In high-pressure design it is desirable to keep parts subjected to high-pressure stress at as low a temperature as possible and to keep parts subjected to high temperature as free of pressure stress as possible. As shown in Figure 2, the T.V.A. converter shell is insulated

from the hot catalyst by a layer of insulation inside the catalyst basket and by a stream of cooling gas passing down between the basket wall and the shell. The shell was originally insulated on the outer surface to prevent excessive loss of heat which the cooling gas picks up in the preheater. However, when the plant was started up, it was found that the upper part of the shell (around the catalyst basket) could not be kept below 450° F., which was considered to be the upper limit from the standpoint of avoiding excess hydrogen attack on the shell. Two measures were taken: (1) the insulation was stripped off the upper section and a shield set around it so that a draft of air passed up between the shield and the shell, and (2) an additional line was installed to provide more cooling gas to the top of the converter. These steps have made it possible to keep the maximum shell temperature down to the present level of 150°-200° F.

One difficulty brought about by cooling the shell is a radial temperature differential in the catalyst. As shown in Figure 5 a differential of up to 150° F. exists between the center and outer thermocouples, a distance of only about 10 in. Structural limitations prevent placing of thermocouples nearer the catalyst basket wall, but it can be assumed that a further drop occurs in the remaining 12 in. from the outer couple to the wall. It appears that better insulation of the catalyst basket or more cooling in the central zone of the catalyst is needed.

Also shown in Figure 5 is a temperature survey taken in February, 1952, after the train 1 converter had been in operation for about eight months with a new charge at the higher rate of about 125 tons/day. As might be expected, the hot point is lower in the catalyst bed and the general temperature level is somewhat higher than in 1943 near the beginning of the first catalyst run, during which the converter operated at a 90-ton rate.

Pressure and space velocity also require adjustment to offset loss of catalyst activity. As is shown in Table 3, both these factors were increased considerably during the 8½-year run. As has been noted, this train (No. 1) was shut down primarily because the plant was being converted to use of natural gas rather than for catalyst failure, although the catalyst clearly had lost much of its activity. Conceivably, the converter could have been operated for a longer period. Pressure could not have been increased much more, since it was close to the design limit of 350 atm., but space velocity could have been increased to about 25,000 at 350 atm., the design capacity of the circulators. The limiting

factor probably would have been control of converter temperature at the higher velocity. At some point, either in velocity increase or in catalyst-activity loss, the hot point would get so low in the converter that any further increase would "blow out" the reaction. At that point the velocity (and therefore production) would have to be decreased to maintain control.

The remaining variables, hydrogen-nitrogen ratio and content of inert constituents and of recirculated ammonia in the circulating gas, present little in the way of operating problems. The recirculated ammonia is ordinarily kept as low as the refrigeration capacity will allow. Inerts are ordinarily not allowed to increase above the range of about 8 to 14% because of the adverse effect on ammonia content of gas leaving the converter, which is kept as high as possible because of limited refrigeration capacity. The hydrogen-nitrogen ratio is kept a little high on the nitrogen side to offset the higher loss of nitrogen dissolved in the product ammonia; it has been found that the nitrogen-hydrogen ratio in the gas flashed off from the ammonia weigh tanks is somewhat higher than in the make-up gas, which indicates a higher solubility for the nitrogen.

Catalyst Performance

The synthesis catalyst used in the T.V.A. plant has been described previously (1). Before reduction, it consists principally of iron oxides (about 97%) with about 1.8% Al_2O_3 and 1.0% K_2O as promoters. The principal problems in maintaining adequate catalyst performance are loss of catalyst activity and nonuniform distribution of gas through the catalyst bed.

One factor that affects activity is the catalyst operating temperature, control of which has been discussed in the previous section. Figure 5 indicates that the center core of catalyst, which was at a higher temperature than outer portions of the catalyst mass, declined in activity at a faster rate during the 8½-year run. This indicates the effect of high temperature, although it should be noted that part of the apparent loss of activity in the center section may have been due to nonuniform distribution of gas flow downward through the catalyst. In interpreting plant data, there is no feasible way of separating the effects of activity loss and gas channeling.

A further problem is the poisoning effect of various impurities that enter the catalyst in the recirculating gas. These include residual water vapor and carbon oxides from the make-up gas; sulfur and phosphorus from lubricating oil; and chlorine from treated water used in the carbon dioxide scrubbers. The

amount of water vapor and carbon oxides entering the converters during normal operation is practically nil. Water vapor is eliminated in the ammonia-cooled condensers, which are placed after the make-up gas inlet (an otherwise undesirable point) for this purpose. Carbon monoxide is removed practically completely in the copper-liquor scrubbing unit, and deposition of ammonium carbonates at various points between the make-up gas inlet and the converter effectively removes carbon dioxide. There is no indication that the small amounts remaining in the gas have had any effect on the catalyst.

The amounts of sulfur, phosphorus, and chlorine entering the converter are difficult to determine. The amounts that accumulate in the catalyst are indicated in Table 4, in which is shown the analysis of catalyst removed from train 1. The most significant increase is in the content of sulfur, which has been found by experiment to be one of the most severe and permanent poisons. The copper content indicated in the table resulted from an accidental influx of copper liquor into the synthesis system in September, 1942, less than two months after the run started. Apparently, this had no adverse effect on the catalyst performance.

Although only a negligible amount of water vapor and carbon oxides enter the catalyst during normal operation, large amounts may enter accidentally for a short time. These compounds are ordinarily regarded as temporary poisons, but small-scale experiments have shown that excessive poisoning results in some permanent damage.

Accidental surges of carbon oxides are caused ordinarily by loss of control in the make-up gas purification section. Water-vapor influx, which is a more frequent and troublesome difficulty, results from steaming out of filters and condensers in the synthesis circuit to remove accumulated carbonate and oil. Failure of valves to close properly, and incomplete removal of condensate prior to putting the unit back in the circuit are the most frequent causes of water getting into the converter. Train 1 has been subjected to accidental surges of carbon oxides and water on a number of occasions. In each case the catalyst activity returned to normal after a temporary drop and there was no apparent

permanent damage. The one major instance of poisoning in train 2 occurred in April, 1949, when water was accidentally swept into the converter. The catalyst activity never returned fully to normal and an accelerated decline began soon after, so that the catalyst had to be replaced less than a year later. It is thought that the catalyst would have recovered from the poisoning if it had been new and at high activity. However, loss of activity was well advanced at the time of the accident and the catalyst apparently had lost much of its ability to recover from poisoning.

The gradual decline in activity in both trains 1 and 2 was likely caused partially by the accumulation of sulfur. Also, microscopic examination of the catalyst showed evidence of pronounced grain growth of iron, which is indicative of crystallization of the iron from its initial amorphous state obtained upon reduction of the original magnetite. The consequent loss of activity appears to be an inevitable result of long-continued catalyst operation.

Another factor that affects catalyst activity is the method used in reducing the catalyst. Experimental work has shown that during reduction the use of high space velocity, low heating rate, and minimum water vapor in the gas phase are conducive to good catalyst activity after reduction. Based on small-scale experiments, an optimum reduction schedule was worked out under which gas is passed over the catalyst (in place in the converter) for 85 hr. at full plant pressure and with a gradual increase in temperature up to the operating level.

In practice it has been found that the piping joints of the starting heater fail at full pressure so reduction is now carried out at about 1500 lb./sq.in. and under a slower schedule that requires about 160 hr. to complete. Operating results indicate that the plant schedule gives as high an activity as that obtained in the experimental work.

Nonuniform flow of gas through the catalyst may result from disintegration of catalyst, improper charging, or from failure of converter parts which allows channeling around the bed. The catalyst taken from the converters was somewhat finer than that charged, but part of the breakage was incurred in removal and so the amount of disintegration within the converter could not be determined. Breakage tests made on new and used catalyst indicated a loss of only about 10% in strength. Moreover, pressure drop through the converter did not increase appreciably over the long runs, never exceeding 10 lb./sq.in. This indicates that catalyst disintegration has not been an important problem. One

Table 4.—Poisons Accumulated in Used Catalyst

DESCRIPTION OF SAMPLE[1]	PER CENT			
	Cl	P	S	Cu
From top of bed	0.05	0.01	0.06	0.04
From bottom of bed	0.01	0.02	0.02	0.01
Composite of bed	0.01	0.01	0.03	0.01
New catalyst	0.02	0.005	0.005	....

[1] Samples of used catalyst from train 1; catalyst was in operation for about 8½ years.

exception which should be noted is the accumulation of fines on top of the catalyst bed, caused possibly by the erosive and grinding effect of the incoming gas. A cone of catalyst particles and dust on top of the bed was found in each of the converters when they were opened, and this likely caused some channeling of gas, at least in the top section. A screen was placed over the catalyst after charging in an attempt to prevent cone formation in the future.

Charging of the catalyst is carried out carefully in order to obtain a bed of uniform density and resistance to gas flow. The catalyst is introduced in small batches with gentle tamping and the basket jarred when full to produce a final settling. A drop in level of about 2 ft. occurred during the catalyst run, presumably because of shrinkage during reduction and some disintegration of catalyst particles.

Some channeling around the catalyst undoubtedly occurred because of failure of the catalyst basket insulation. This apparently disintegrated and passed on into the gas stream, leaving an open annular space between the basket wall and the insulation cover. Seal welds had also broken loose, which allowed gas to pass into and down through the annular space.

Loss of Process Material

Loss of process material because of leaks from various closures, compressor packing, and valves is an important problem in high-pressure design and operation. In the T.V.A. plant the principal closure, that for the converter heads, is a Bridgeman-type gasket seal (Fig. 2). The gasket is of copper and is forced against the converter wall by the floating head, which is forced upward against the sloping gasket by the internal converter pressure. This closure has maintained a perfect seal; the only difficulty is that in seven or eight years of operation the gasket bonds tightly to the steel surrounding it and must be drilled out.

All other high-pressure joints and closures such as for tubing, separators, and filters, are of the lens-ring type. The sizes range up to 20 in. for the filters. The lens rings for tubing joints are made from a section of tubing approximately one wall thickness in length. Both the ring and the tube ends are polished smooth after machining. The number of tubing joints is reduced by welding, wherever possible.

Although these joints require careful machining, and are therefore costly, they give excellent service and are simple to assemble and dismantle. One exception should be noted, however; at the high temperature and pressure encountered in the starting heater the joints fail quickly. It has been found necessary to restrict pressure in the heater to 2000 lb./sq.in. and temperature of heated gas to 850° F. Under these conditions the joints have given reasonably good service. No other joint in the synthesis circuit is exposed to a temperature of more than 375° F.

There has been one serious leak during the history of the plant. This came from a crack in the bottom forging of the train 2 converter which developed during the initial start-up in 1942. It would normally be expected that such a leak would increase because of the erosive action of the escaping gas. No increase was noted, however, and since some purging is necessary anyway, it was considered preferable to continue operation rather than to shut down for repair. When the train was finally shut down in 1950, the leak was still of about the same magnitude as at the beginning. In Figure 6, the crack is made clearly visible by means of powder collected by Magnaflux. Figure 7 is a view showing the forging after sufficient steel had been removed to eliminate the crack, and Figure 8 shows the result after the forging had been built back up with weld metal and a new gasket seat machined.

Other leaks are encountered around packing in valves, circulators, and compressors, and these are minimized by choice of packing and by careful maintenance.

One of the major factors affecting process efficiency from the standpoint of conservation of raw material is the necessity for purging part of the synthesis gas to keep down the content of inert methane and argon in the synthesis circuit. Leaks serve as part of the purge; in fact, during the first run of train 2 leakage from the cracked converter, forging was sufficient for the entire purge. Normally, however, gas is bled from a point just before the circulators, which is the point of highest inert concentration.

An additional amount of gas dissolves in the ammonia produced and is subsequently lost when it is flashed off in the weigh tanks. However, this is quite an

Fig. 6. Bottom forging from ammonia synthesis converter showing crack as revealed by Magnaflux.

Fig. 7. Bottom forging from ammonia synthesis converter after boring and drilling to remove crack.

efficient purge since methane and argon dissolve in the ammonia in relatively higher proportions than do the hydrogen and nitrogen (Fig. 1).

The gas bled from the system is passed through an ammonia-recovery system where the amount recovered is approximately 5% of the total production under current operating conditions. The total amount of hydrogen lost is normally between 3 and 4% of that in the make-up gas.

Materials of Construction

Materials of construction used in ammonia synthesis present an operating problem since, after the designer has selected and installed the best available materials for the severe conditions involved, the operator must select process conditions that will give as long a life to the materials as is consistent with economical and dependable operation.

Problems involving materials of construction have occurred in the T.V.A. synthesis units only in the converters. The converter proper (Fig. 2)

dence of hydrogen attack at any point, either on the alloy liner or on the forgings.

A further problem anticipated in design of the converter shell was diffusion of hydrogen through the inner layer and a resulting build-up of pressure between layers. To relieve any such pressure "tell-tale" holes were bored at intervals through the outer layers up to the liner. However, there has been no indication of leakage in operation.

The problem of materials selection is considerably different for the internal parts of the converter, where the problem is high temperature rather than pressure stress. Hydrogen attack is minimized by use of a low-carbon steel with chromium added to stabilize carbides. As to attack by nitrogen, apparently almost any steel will nitride eventually at high temperatures, although the austenitic chromium-nickel types are reported to be the most resistant (4). In view of this fact, the approach in designing the high-temperature portion of the T.V.A. converter was to keep the design as simple and rugged as possible, with a

the catalyst basket is lined with a layer of insulation which is held in place with a tight-fitting steel cover.

Examination of these parts at the end of the runs showed no identifiable hydrogen attack, but the expected nitriding had taken place to an advanced degree in the exchanger and catalyst cooling tubes (3). Most of these, however, were apparently still functioning properly within the converter, even though they were severely nitrided; a test made on the train 2 exchanger showed that only 52 of the 1086 tubes were leaking and there was no indication of failure of any of the cooling tubes.

The catalyst basket shells were in good condition except for a cracked flange on train 2. This basket was further injured in disassembly and as a result was scrapped; the train 1 shell, however, probably will be reused. The exchanger shells were also in good shape except for a cracked weld in train 1 (Fig. 9); the train 2 exchanger was retubed and replaced in the converter in 1950, and the other will probably be reused also.
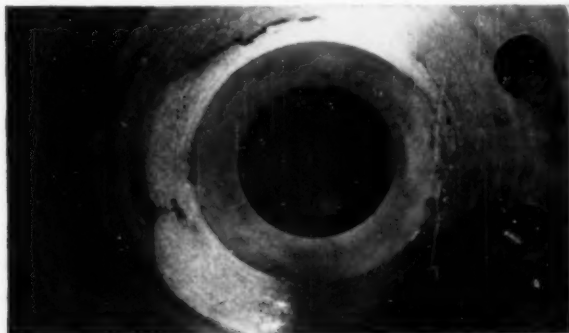
Fig. 8. Bottom forging from ammonia synthesis converter after repair of crack.
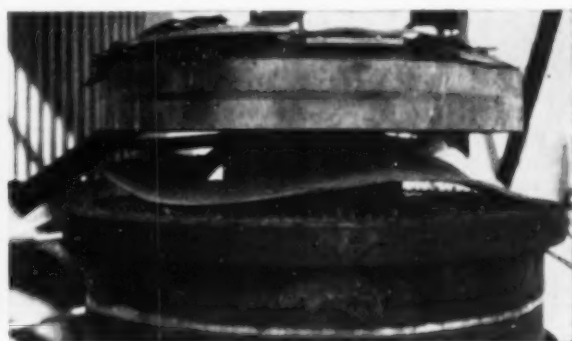
Fig. 9. Gas distributor and heat exchanger from ammonia synthesis converter after 8½ years' service.

is a 48-in. I.D. cylindrical vessel. The shell is of multilayer construction with successive layers of carbon steel plate welded on both longitudinal and lateral seams, plus an inner ½-in. layer of chrome steel to give a total wall thickness of 8½ in. Machined forgings are welded to the ends of the cylinder to form closures.

The main problem in selection of materials throughout the converter is protection against attack by hydrogen and nitrogen. For those parts exposed to high pressure, reliance is placed more on keeping the parts at a relatively low temperature than on use of special materials, although the chrome steel liner for the shell is an added protection for the area adjacent to the hottest parts in the converter.

The operating problems involved in keeping the shell cool have been described. The cooling methods used have been quite satisfactory, as shown by the condition of the shells and forgings when the converters were opened after 7½ and 8½ years' service. There was no evi-

minimum of joints and welds where stresses might cause failure after eventual nitriding. In addition, care has been taken in operation to avoid thermal shock on the interior parts by making any necessary changes in temperature at a slow, even rate. Through careful operation it has been possible also to avoid many of these changes by keeping the converter hot during temporary interruptions in production. As a result the converters have given good service, as is indicated by the long period of operation without internal maintenance.

Most of the internal parts, such as catalyst cooling and heat exchanger tubes, and the catalyst basket and exchanger shells, are made of A.I.S.I. Type 501 steel. The gas distributor between the exchanger and basket, a point of especial stress, is of Type 410. In addition the distributor is covered and

The gas distributor, the upper exchanger tube sheet, and the catalyst grid support were in fairly good condition, although these parts are subjected to the most severe temperature conditions in the converter. All are reusable except that damage to the threads in the distributor during removal of the cooling tubes probably will make it desirable to replace this part.

The principal points of damage (other than the tubes) were the catalyst basket insulation and the insulation covers for the distributor and catalyst basket. Figure 9, a view of the distributor and top of the exchanger from the train 1 converter, shows that the insulation cover was in bad condition. Figure 10 shows the poor condition of the insulation cover for the train 2 catalyst basket. All five sections of this cover were broken loose from their supporting

ribs on the basket shell, and the insulation behind the cover was either shrunk badly or missing.

In the recent repair and replacement of the converter internals, no changes were made except in the basket insulation and the insulation covers. The basket insulation was changed to a type that is reported to be good up to 1500° F., and covers were changed to Type 304 steel. Materials used for all other parts had given satisfactory service and therefore no changes were considered necessary. The insulation covers are of relatively thin sheet material and were subject to warping and buckling, which probably accounted for their failure.

Operating Hazards

A number of operating hazards are involved in ammonia synthesis, both because of the high pressure used and because of the combustible and toxic gases handled. Training of the operating personnel is especially important because serious accidents could easily result from incorrect action in an emergency. One point that should be stressed is that changes in operating conditions must be made slowly and carefully.

One of the principal hazards encountered has been the blowing out of sight glasses in control of the liquid level in the ammonia separators. The position of the separators in the synthesis circuit requires that a uniform level be maintained in the vessel. Should the level rise too high, liquid ammonia would enter the converter or the circulators; if it should drop too low, gas would escape into the weigh tanks.

The first method employed to determine the liquid level was by use of try-cocks, and the level was maintained by manual operation of a valve in the pressure breakdown assembly. This method required constant attention by an operator, and maintenance of the try-cocks was excessive. In an effort to improve this operation, glass sight columns were installed, but the ammonia draw-off was still manually controlled. After the blowing out of several sight glasses, one of which was responsible for a fire, automatic level controls were installed. Each separator was equipped with a float assembly (float-displacement type, torque tube and pilot) in place of the sight column, and a diaphragm motor valve in place of the manual level control valve. This arrangement now permits close control of the liquid level in the separator, completely eliminates any explosion hazards, and indicates the liquid level at a remote point. In addition it has allowed a considerable reduction both in operating and maintenance labor. The try-cock and sight-glass systems are shown in Figure 11 and the automatic control in Figure 12.

Another source of hazard is the accumulation of explosive and toxic gases in the operating area. Measures taken to

Fig. 10. Catalyst basket liner after 7½ years' service.

Table 5.—Ammonia Plant Process Interruptions
(1943 through 1951)

SOURCE	DAYS	% OF CAPACITY PRODUCTION
Compressors	32	1.0
High-pressure section	28[1]	0.9
Other sections of ammonia plant	27	0.8
Plant total	87	2.7
Causes extraneous to ammonia plant	63[2]	1.9
Total	150	4.6

[1] Includes one catalyst change in each converter.
[2] For causes such as shortage of process steam or change-over to natural gas.

prevent or counteract this are as follows:

1. Relief valves and rupture discs: All discharge piping is routed outside of the building, clear of the operating area.

2. Welding in plant area: Equipment must be completely isolated, purged, and properly blanked; atmosphere must be analyzed for combustible gases; and high-pressure steam must be available to control any fires.

3. Smoking: Not allowed in equipment area.

4. Oxygen in equipment and piping: No oxygen is allowed in any vessel or piping before introduction of synthesis gas. Equipment is purged at slightly above atmospheric pressure with steam or an inert gas.

5. Piping vibration: In some cases vibration of pipe against the hanger has worn the pipe wall thin. Change of the hanger and anchor design has eliminated this difficulty completely.

Summary

Although process operation at high pressure introduces problems that are not encountered at normal pressures, modern design has progressed to the extent that a high-pressure process such as ammonia synthesis is dependable and efficient in operation. Operating experience in the T.V.A. plant is summarized

in Table 5, in which all interruptions in production are accumulated and represented as days of plant down time for the nine-year period from 1943 through 1951.

The operating experience has indicated a number of conclusions which may be drawn regarding the ammonia synthesis process used in the T.V.A. plant:

1. Compressor maintenance is the principal cause for process down time, but interruptions have not been of such magnitude as to make a spare compressor economical.

2. Operation of high-pressure valves is improved by Teflon packing.

3. Further improvement in means for removing solid and liquid impurities from synthesis gas is needed.

4. Temperature of the catalyst is the most important operating variable; every effort should be made to keep the temperature at a minimum. Other variables must be adjusted to take care of continuing loss of catalyst activity.

5. Sulfur accumulation from lubricating oil is possibly the chief cause of loss of catalytic activity, although a number of other factors undoubtedly play a part.

6. The Bridgeman type of closure for large vessels gives an excellent seal, but the gasket tends
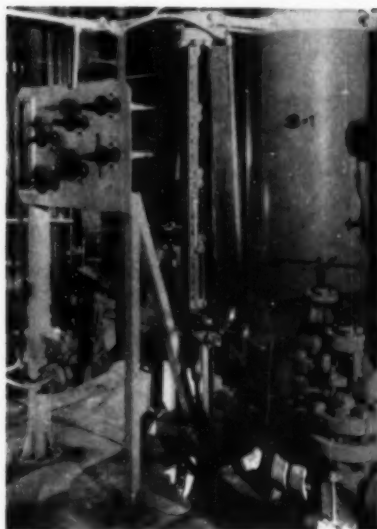
Fig. 11. Arrangements for liquid level control in ammonia separators. Try-cock panel at left; sight-glass installation in center; drawdown valve in right foreground.

to stick. Lens-ring joints for piping and small vessels give good service except at high temperature and pressure.

7. Simplicity of construction appears to be the most important requirement in design of the synthesis converter.

Acknowledgments

The authors wish to acknowledge the contribution of the entire ammonia plant operating staff, from whose experience most of the material in this paper is drawn. Especial credit is due E. J. O'Brien and J. L. Snyder of the operating staff; to C. O. Brown and J. G. Dely, consulting chemical engineers; to the staff of Stone and Webster Engineering Corp., and to the T.V.A. design staff.

Various experimental data listed resulted from the work of various members of the T.V.A. research, development, and plant chemical control staffs.

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Discussion

C. S. Carlson (Standard Oil Dev. Co., P. O. Box 51, Linden, N. J.): I notice in your slides a complete absence of barricades. What protective devices do you provide for your operating personnel?

A. V. Slack: We do not consider that there is sufficient hazard in operation of high-pressure equipment of this type to warrant the use of barricades. There is some possibility of explosion, of course, because of the presence of hydrogen under high pressure and the possibility of leakage into the operating area. However, regulations concerning smoking and welding in the area appear to be an adequate safeguard for the degree of hazard involved.

J. W. Bulls (Commercial Solvents Corp., Terre Haute, Ind.): Would you care to comment on the reasons for the decrease in your catalyst activity?

A. V. Slack: One cause for loss of activity in any iron-base ammonia catalyst is crystallization of the iron from its original amorphous state, a condition which gradually develops on extended use. A further cause in the T.V.A. operation was the accumulation of poisons such as sulfur and chlorine in the catalyst over the operating life of 8½ years. Sulfur has been the principal contaminant; it could have come from the water gas formerly used as a source of hydrogen, but a more likely possibility is that it came from oil used in the compressors and circulators. A low-sulfur oil is used in this equipment, and care is exercised in operating the filters, but nevertheless some sulfur gets through to the catalyst. It would be difficult to determine how much of the loss of activity was due to poisoning and how much to age and resulting crystallization of iron, but undoubtedly both contributed.

C. E. Alleman (Phillips Chemical Co., Duma, Tex.): How did you determine the decrease in catalyst activity? In other words, what measurements did you use to arrive at the conclusion that it had decreased?

A. V. Slack: We have a small converter hooked into the operating circuit. Gas from the synthesis circuit can be tapped off and passed through a sample of catalyst placed in this converter, thereby duplicating synthesis conditions. After a charge of catalyst is taken out of service, we place a sample in the test converter and determine its activity in terms of conversion efficiency. We then test a sample of unused catalyst from the original batch and thus get an accurate comparison under actual plant operating conditions.

R. A. Ghelardi: (Merck & Co., Inc., Danville, Pa.): I didn't notice any safety devices on your high-pressure system. What are you using to protect yourself against overpressure—rupture discs, or pop valves?

H. E. Maune: We have relief valves on the last stage of the compressors. No rupture disks are used on high-pressure equipment; however, they are used on the low-pressure
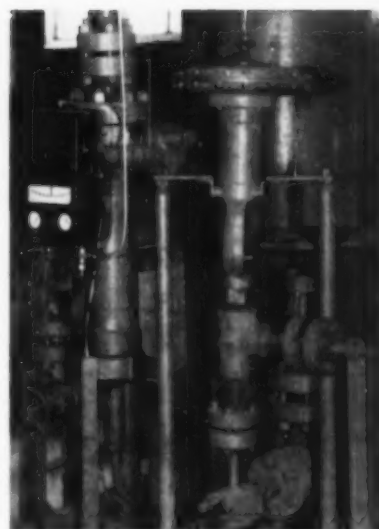
Fig. 12. Installation for automatic level control in ammonia separator.

side of equipment such as the separators, to provide against accidental development of high pressure. We have had no particular operating problems with relief valves or rupture disks.

L. L. Baldwin: (Shell Chemical Corp., New York, N. Y.): The heat exchanger in your high-pressure ammonia converter is shown in your drawing to support the weight of the catalyst and catalyst basket. What is the construction material of your inter-exchanger tubes, and have you experienced nitriding of the tubes at the upper or hot tube sheet?

A. V. Slack: The exchanger tubes are made of low carbon steel containing about 5% chromium. The tubes were nitrided but not as much as the catalyst cooling tubes, since the latter are exposed to a higher temperature. Depth of nitriding was much greater on the cooling tubes. However, some of the exchanger tubes developed leaks in service whereas the cooling tubes did not, mainly because the exchanger tube walls are much thinner.

C. F. Smith (Allied Chemical & Dye Corp., Marcus Hook, Pa.): I am interested in the speed with which you can change the valves in the compressor. Can you tell us whose compressors they are and if there are special techniques involved in these rapid changes?

A. V. Slack: The original installation consisted of four Cooper-Bessemer compressors of the 100 rev./min., tandem-cylinder type. Two new Clark compressors of the horizonally opposed type were installed in 1951. There is no particular trick in changing valves rapidly. It merely requires well-trained maintenance and operating personnel and a well-organized routine for making the change.

A Continuous Countercurrent Ion-Exchange Unit

Ralph H. McCormack and John F. Howard

University of Detroit, Detroit, Michigan

The transfer of valuable solutes from dilute solutions to relatively concentrated solutions by continuous countercurrent ion exchange is an attractive possibility compared to concentration of such solutions by evaporation. Processes, using various techniques to effect this transfer have been patented (1, 2, 4, 5, 7) and it has been shown (6) how a fluidizing process can be carried out in producing concentrated copper solutions from dilute copper solutions. The fluidized resin equipment, however, requires circulation of a resin stream, and the present work was undertaken to develop a countercurrent ion-exchange unit which would not require such circulation.

The apparatus shown in Figure 1, was the result of this project. This unit comprising essentially an endless bed of ion-exchange resin enclosed in a porous flexible casing, moved slowly through a series of three tubular liquid reservoirs. The first reservoir was the feed exchange section in which the ion to be recovered was removed from the feed solution by the ion-exchange bed. The next reservoir was the regenerating section in which the exchanged ion was recovered from the ion-exchange bed as product solution. The last reservoir was a rinsing section to remove adherent

Mr. Howard is presently associated with Du Pont Co., Wilmington, Del.

regenerating acid from the resin bed. These reservoirs were sized to fit as closely as possible around the bed, to prevent solution by-passing the resin, and shaped to the sag of the resin bed between the drive wheels and between the drive wheels and the idler so as to minimize the drag of the bed against the tube walls.

Pyrex glass liquid reservoirs were used first with this unit but were replaced with polyethylene tubing. Two resin beds were tested, one containing a section of zeo-karb and a section of Nalcite HCR, the other entirely of zeo-karb. Both beds were 30 ft. long with the resin enclosed in Lumite filter fabric, sewed with Saran thread, 700/3, and tied every 6 in. to prevent the resin from shifting. The drive pulley and idlers were made of water-resistant plywood grooved to fit the bed with the grooves lined with rubber. The resin bed was moved through the system at a rate of 6 ft./hr. by a drive from a small motor. Flow rates were controlled by reference to capillary manometers and measured by weighing. Flow was from the overhead solution tanks through the flowmeters, through the three sections of the unit, and out to reservoirs placed beneath the apparatus. Operation was followed by periodic determination of the pH of the streams; a Beckman model H-2 pH meter was used.

The operation of this unit was demonstrated with copper sulfate feeds of 0.34-0.36% Cu by weight, leading to con-

centrated solutions of 2.7-3.2% Cu on regeneration with 2.0 N sulfuric acid with 95% recovery of the copper in the concentrated solution. Another copper sulfate feed of 0.11% Cu produced a 0.95% Cu solution when regenerated with 1.02 N sulfuric acid with complete recovery of the exchanged copper in the concentrated solution. Copper was determined in all cases by the iodometric method described by Hall (3).

This continuous countercurrent ion exchanger was developed for the concentration of any exchangeable ion by transfer from a dilute solution to a relatively concentrated solution. Copper was used as a convenient ion for demonstrating the performance of the unit.

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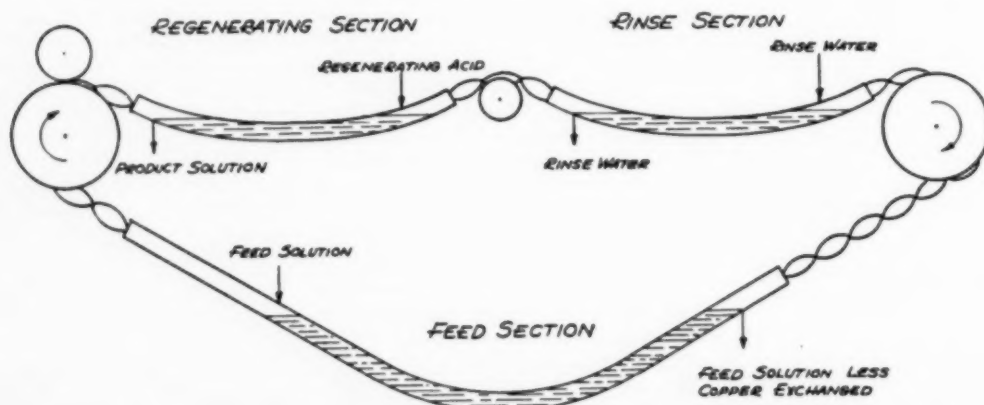
Fig. 1. Continuous countercurrent ion-exchange unit.

The Effect of Packing Size and Column Diameter

on Mass Transfer in Liquid-Liquid Extraction

Irving Leibson and R. B. Beckmann Carnegie Institute of Technology, Pittsburgh, Pennsylvania

During recent years the widespread interest in and increasing use of liquid-liquid extraction in industrial processes has made the need for more basic design information apparent. This paper is concerned primarily with the effect of packing size and column diameter on the rate of mass transfer in continuous countercurrent liquid-liquid extraction. The common practice in the design of packed towers has been to use a ratio of column diameter to packing size of not less than 8.0, as previously established for gas absorption (*18* and *21*). It would be economically advantageous if the variables of column diameter and packing size could be balanced against the number of towers and the tower height necessary to secure a given degree of separation at a required throughput.

There have been a few experimental investigations of mass transfer in continuous countercurrent liquid-liquid extraction reported in the literature in which either column diameter or packing size has been varied. For the most part, these studies have been confined to small-diameter columns using fine packings and, in some cases, larger industrial-size packings. Varteressian and Fenske (*22*) conducted a performance study on a 0.55-in. I.D. packed column for the transfer of ethyl alcohol from benzene to water. No. 19 single-link aluminum and copper chain packings and No. 26 B. & S. gauge nickel wire made into 5/32-in. ring packing were used. They found that a higher extraction efficiency was obtained with the smaller packing size and attributed this to the smaller drop size of the discontinuous phase in this packing, which

This investigation is a study of the effect of packing size and column diameter upon mass transfer in countercurrent liquid-liquid extraction. Six different packings were used: ¼-, ⅜-, ½-, ⅝-, ¾-, and 1-in. nonporous, unglazed porcelain Raschig rings. Three extraction columns, 3, 4, and 6 in. I.D., were used in the experimental work. The mass transfer of diethylamine from a continuous water phase to a dispersed toluene phase was investigated.

Photographs of the dispersed-phase droplet behavior in the packing during the column operation revealed that the droplet size of the dispersed phase is (*a*) independent of packing type and size for packings ½ in. or larger and (*b*) greater for ¼-in. rings than for any of the larger sizes of packing.

The photographic analysis permitted an explanation of the mass-transfer results in terms of changes in the interfacial contact area due to the effect of the packing size on the dispersed-phase droplet size and on the holdup of the dispersed phase. A correlation of the effect of packing size on the individual film height of a transfer unit was developed for packing sizes ½ in. or larger.

resulted in increased interfacial contact area for a given throughput.

Morello and Beckmann (*14*) studied the extraction of diethylamine from water with toluene in a 1⅝₆-in. I.D. column using 4-mm. glass beads. Smythe (*20*) made a study of the same system using 4- and 6-mm. glass beads. The results indicated a 20 to 30% increase in the over-all mass-transfer coefficient for the larger packing. These results contradict the work of Varteressian and Fenske. It is possible that this difference in effects is connected with differences in the shape of the void spaces of the solid bead packing as compared with the more open chain- and ring-type packings.

Rushton (*17*) reported the extraction of oil by nitrobenzene in a 2⅝₆-in. I.D. column in which packing size was varied from 0.16 to 0.50 in. for glass rings. The other packings used were ⅜-in. unglazed stoneware and porcelain Raschig rings, ¼- and ½-in. carbon rings, and ½- and 1-in. unglazed stoneware Berl saddles. Since most of the data were taken with a graded packing, using different sizes in the various sections of the tower, only qualitative evidence was obtained of increased column efficiency with smaller packings. Comings and Briggs (*6*) studied four wetted-wall

and four packed columns varying from 0.47 to 7.45 in. I.D. and from 4.6 to 46.5 in. in height, using a water-benzene system with benzoic acid, aniline, and acetic acid as the solutes. Packings used were ½- and 1-in. Berl saddles and ½-, ¾-, and 1-in. Raschig rings. Most of the data were taken with water as the dispersed phase. The results obtained using a packed height of 4.6 in. with packing sizes varying from ½ to 1 in. can hardly be expected to be representative of a randomly packed bed. The data on the effect of packing size were too meager to afford more than a qualitative indication of a higher over-all mass-transfer coefficient with smaller packings.

Sherwood, Evans, and Longcor (*19*) studied the mass transfer of acetic acid from water to benzene and from water to methyl isobutyl ketone in a 3.55-in. I.D. column. The column was operated both as a spray tower and as a packed tower with ½- and 1-in. carbon rings and ½-in. stoneware Berl saddles. The ratio of column diameter to packing size was 7.0 for the ½-in. packing and only 3.5 for the 1-in. packing. For both systems, the ½-in. packing gave the greatest over-all mass-transfer coefficients. The 1-in. packing, compared to spray-column operation, served only to in-

crease the holdup of the discontinuous phase. Several points were taken with two different packed-bed heights, but the data were too meager to draw any definite conclusions concerning the effect of tower height on the over-all mass-transfer coefficient.

Claffey, Badgett, Skalamera, and Phillips (4) investigated nicotine extraction from water with kerosene in a 1¾-in. I.D. column using ¼- and ½-in. ceramic Berl saddles and ⅜- and ½-in. ceramic Raschig rings. The ratio of tower diameter to packing size was 7.5 for ¼-in. saddles, 5.0 for ⅜-in. rings, and 3.8 for ½-in. rings and saddles. On the basis of data at 68° C., the ¼-in. Berl saddles gave the highest over-all mass-transfer coefficients followed by ⅜-in. rings, ½-in. saddles, and ½-in. rings respectively. There was a tendency for the saddle packing to settle, decreasing the void fraction. In the case of ¼-in. saddles, this resulted in lower mass-transfer coefficients. It is possible that this condition may have stemmed from an increase in drop size with the smaller void spaces of the settled ¼-in. saddles (as pointed out by Lewis, Jones, and Pratt (11) for ⅜- and ¼-in. ring packings).

Laddha and Smith (10) studied two binary systems, 3-pentanol-water and isobutyraldehyde-water, in a 2-in. I.D. column operated as a spray tower and as a packed tower with ¼- and ⅜-in. Raschig rings. The purpose of the study was to evaluate H_t (height of a transfer unit) values for the individual films by the method of Colburn and Welsh (5). The authors reported that the H_t of the discontinuous phase was nearly identical for the three types of column operation. However, from a careful consideration of the data, it appears that in almost all cases, spray-tower operation is less efficient than packed-tower operation. The effect of varying the packing size on the H_t of the continuous phase was very slight. There was a 10 to 20% decrease of $(H_t)_c$ from the ⅜- to the ¼-in. rings for isobutyraldehyde dispersed in the isobutyraldehyde-water system and for water dispersed in the 3-pentanol-water system. There was even less difference between the ⅜- and ¼-in. rings for 3-pentanol dispersed in the 3-pentanol-water system.

No other fundamental studies of mass transfer in liquid-liquid extraction in which either column diameter or packing size has been varied have been found in the literature. The work with fine packings is contradictory with respect to the effect of varying packing size. In the majority of the work in which industrial packings were used, the ratio of column diameter to packing size was considerably less than 8. As a result, the void fractions of the larger packings
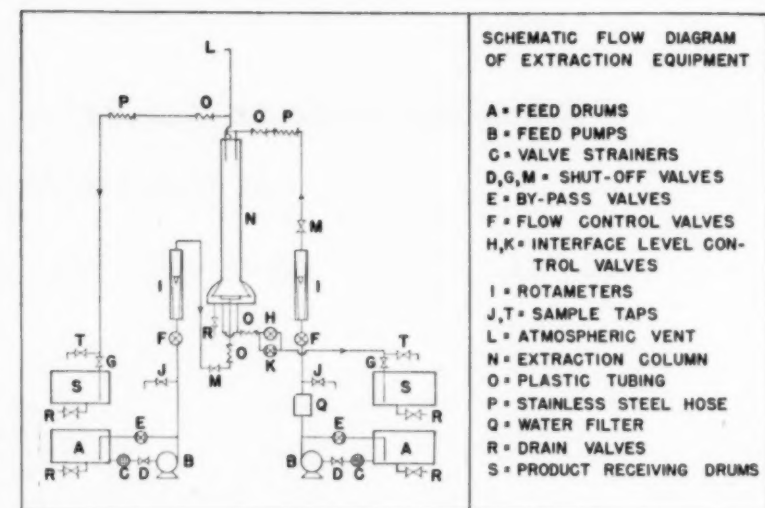
must have been greater than normal. Nevertheless, the qualitative result of the previously reported work is that the over-all mass-transfer coefficient, Ka, decreases with increase in packing size for industrial packings.

Apparatus

A schematic flow diagram of the extraction equipment is presented in Figure 1. The material of construction for all lines, valves, valve strainers, and column parts (unless otherwise stated) was type-304 stainless steel. All lines were ½-in. pipe with the exception of short sections of ¼-in. pipe on the rotameter control panel, the column vent line, and in the column exit lines at the raffinate and extract receiver drums. The feed drums and the raffinate and extract receiver drums were standard 55-gal. drums, black iron for toluene and galvanized iron for water. The feed pumps were Worthington type 1-DN-2 centrifugal, close-coupled pumps.

Flow rates in the system were regulated by ¼-in. needle valves with forty micrometer threads to the inch. The flow of the toluene and the water streams entering the column were measured by rotameters covering a flow range from 0.01 to 0.89 gal. of water/min. Two 3-ft. lengths of stainless-steel hose were employed to provide flexibility in the elevation of the lines at the top of the column. This arrangement permitted the variation of column height without changing the flow system. The top of the extraction column was vented. The level of the two-phase interface was regulated by means of a ½-in. globe valve for coarse adjustments paralleled by a ¼-in. needle valve for fine control.

The extraction column is shown in detail in Figure 2. Three extraction columns, 3, 4, and 6 in. I.D., were used in the experimental work. The structural features of each of the columns used were identical. The pertinent dimensions are given in Figure 2. The extraction column

consisted of several 2-ft. flanged sections of Pyrex-glass pipe and a Pyrex bell section fitted with a 20-gauge stainless-steel cone. The aqueous phase entered the column through two sections of ¼-in. pipe which ended flush with the top of the column packing. The aqueous raffinate was removed from the column through two ½-in. pipes threaded into the base flange of the column.

Toluene entered the column through the dispersed-phase distributor. The toluene was dispersed by a number of sharp-edged nozzles (0.106 in. I.D., 0.014 in. wall thickness, and 1¼ in. long) made from stainless-steel hypodermic tubing. These nozzles were spaced in concentric circles with a minimum circumferential distance between nozzles of ½ in., and they were soldered to the top of the dispersed-phase distributor. The number of nozzles used on each of the dispersed-phase distributors for the various sizes of the columns was maintained approximately proportional to the column cross-sectional area. The number of nozzles per distributor for each column is given in Figure 2. The dispersed-phase distributor was flush with the bottom of the packing with the nozzles imbedded 1¼ in. into the packing. Thus, the cross-sectional area of the annular space between the dispersed-phase distributor and the stainless-steel cone at this point was equal to the cross-sectional area of the column as recommended by Blanding and Elgin (2).

The drops of toluene rose through the column packing into the separating section, where they coalesced into the layer of toluene extract above the two-phase interface. The toluene extract overflowed from the column through a ½-in. pipe threaded into the top flange of the column.

The support screen for the column packing was type-304 stainless steel, 3¾ mesh with 0.054-in. wire and openings 0.213 in. square. This screen was held in place by the dispersed-phase distributor and by four ½-in. stainless-steel rods which were threaded into the base flange of the column. The bell section was sealed to the

Fig. 1. Schematic flow diagram of extraction equipment.

SCHEMATIC FLOW DIAGRAM OF EXTRACTION EQUIPMENT

A = FEED DRUMS
B = FEED PUMPS
C = VALVE STRAINERS
D,G,M = SHUT-OFF VALVES
E = BY-PASS VALVES
F = FLOW CONTROL VALVES
H,K = INTERFACE LEVEL CONTROL VALVES
I = ROTAMETERS
J,T = SAMPLE TAPS
L = ATMOSPHERIC VENT
N = EXTRACTION COLUMN
O = PLASTIC TUBING
P = STAINLESS STEEL HOSE
Q = WATER FILTER
R = DRAIN VALVES
S = PRODUCT RECEIVING DRUMS

EXTRACTION COLUMN DESIGN

DIMENSIONS ARE IN INCHES			NUMBER
①	②	③	OF NOZZLES
3.0	4.9	9.0	27
4.0	6.5	12.0	48
6.0	8.4	18.0	105

A = PYREX GLASS PIPE
B = PYREX BELL SECTION
C = STAINLESS STEEL CONICAL SECTION
D = WATER INLET TUBES
E = PACKING GLANDS
F = TOP FLANGE
G = WATER EXIT LINES
H = BOTTOM FLANGE
I = DISPERSED PHASE DISTRIBUTOR
J = PACKING GLAND
K = TOLUENE EXIT LINE
L = PACKING SUPPORT SCREEN
M = SUPPORT RODS
N = RING FLANGE
O = THERMOMETERS
P = CORK GASKETS

◄ Fig. 2. Details for extraction column design.

of 98 wt. % minimum purity. Accurate values for the equilibrium-distribution constant for this system as a function of temperature and solute concentration were available from the work of Morello (13) and Wehn and Franke (23). From the mutual solubility data of Wehn and Franke it is evident that at concentrations of less than 10 wt. % diethylamine for either the water or the toluene phase, the solubility of one phase in the other is very limited. Thus, for this system, it is possible to neglect the presence of

base flange by means of a ring flange machined from a ½-in. aluminum sheet. The details of this joint are shown clearly in Figure 2. The base flange of the extraction column was bolted firmly to the structural-steel stand through a vibration-insulating material.

All gaskets used between the glass sections of the column were fabricated from ⅛-in. Johns Manville 84-S asbestos sheet. The gasket material used between the glass section of the column and the top or bottom flange was ¼-in. hard-cork sheet.

Column Packings

Six different packings were used: ¼-, ⅜-, ½-, ⅝-, ¾-, and 1-in. nonporous unglazed porcelain Raschig rings. The actual dimensions of these packings are summarized in Table 3. The procedure adopted for the measurement of the superficial area, a_p, and the void fraction, F, of the packing was as follows: The number of pieces of packing dumped into the column was determined by dividing the total weight of the packing by the weight per piece of packing. The volume of the column to be packed was accurately measured. Since the surface area and the volume per piece of packing were known, the superficial area, a_p, and the dry void fraction, F, of the packing could be calculated. These values are reported in Table 2. For the most part, the values for the void fraction of the packing by drainage agreed reasonably well with the calculated dry-void fraction.

The column was packed by dropping several pieces of packing at a time into the column filled with water. The column was not disturbed in any way while the packing was being added. No attempt was made to obtain the densest possible arrangement of the packing. No settling of the packed bed during operation was observed for any of the packings used. The dimensions of the openings of the packing-

support screen were sufficiently small (0.213 in. square) to retain any size of packing used.

Liquid-Liquid System

The toluene-water system was studied with diethylamine as the solute. The materials employed were nitration-grade toluene, distilled water, and diethylamine

the third component in dealing with either phase alone; e.g., in the water phase only diethylamine need be considered. This validates the simplifying assumption that the volume of each liquid phase throughout the column is constant for dilute solutions. For this system, the assumption that the equili-

½-in. rings 3-in. column $V_W = 12.24$ $V_T = 9.52$ 3a	¼-in. rings 6-in. column $V_W = 10.02$ $V_T = 5.18$ 3b	¼-in. rings 6-in. column $V_W = 20.38$ $V_T = 5.18$ 3c	⅜-in. rings 6-in. column $V_W = 7.86$ $V_T = 8.94$ 3d
⅜-in. rings 6-in. column $V_W = 5.78$ $V_T = 14.20$ 3e	⅜-in. rings 6-in. column $V_W = 10.02$ $V_T = 5.18$ 3f	⅜-in. rings 6-in. column $V_W = 30.20$ $V_T = 5.18$ 3g	½-in. rings 3-in. column $V_W = 51.80$ $V_T = 20.72$ 3h

Fig. 3. Photographs of extraction column operation.

TABLE 1.—6-IN. I.D. COLUMN, ½-IN. RINGS

| Run | Concentration, lb. moles/cu.ft. | | | | Column Temp. °C. | | $H = \dfrac{c_W}{c_T}$ | | Material balance, % Error | Flow rate, cu.ft./(hr.)(sq.ft.) | |
| | Toluene | | Water | | | | | | | Toluene | Water |
	Inlet c_{T_2}	Exit C_{T_1}	Inlet c_{W_1}	Exit c_{W_2}	T_1	T_2	H_1	H_2		V_T	V_W
6-3-1	0	0.00602	0.01509	0.00995	30.2	28.7	1.188	1.241	3.1	5.18	7.86
2	0	0.00620	0.01509	0.01053	30.1	28.7	1.188	1.241	3.1	5.18	7.86
3	0	0.00512	0.01509	0.00821	30.2	29.6	1.183	1.275	-6.7	8.94	5.80
4	0	0.00631	0.01509	0.00938	30.2	29.9	1.183	1.245	16.1	5.18	10.02
5	0	0.00698	0.01574	0.01100	29.3	28.5	1.223	1.299	6.8	3.05	5.80
6	0	0.00743	0.01574	0.01241	29.9	29.4	1.195	1.239	2.8	3.05	7.86
7	0	0.00797	0.01574	0.01274	30.5	29.9	1.167	1.212	3.7	3.05	10.02
8	0	0.00831	0.01574	0.01372	30.6	30.5	1.162	1.177	-0.3	3.05	12.27
9	0	0.00872	0.01574	0.01413	30.7	30.6	1.158	1.170	-0.9	3.05	15.12
10	0	0.00860	0.01574	0.01450	30.8	30.8	1.153	1.159	-0.2	3.05	20.40
11	0	0.00778	0.01579	0.01372	28.0	27.0	1.289	1.357	-3.8	5.18	15.12
12	0	0.00727	0.01579	0.01212	28.2	27.5	1.278	1.343	-4.0	8.94	15.12
13	0	0.00672	0.01579	0.01055	28.1	27.7	1.283	1.348	-2.6	12.72	15.12
14	0	0.00606	0.01579	0.00836	27.9	27.7	1.294	1.377	-0.7	18.84	15.12
15	0	0.00729	0.01579	0.01253	27.9	27.7	1.294	1.329	3.2	8.94	23.70
16	0	0.00762	0.01579	0.01446	27.9	27.8	1.294	1.309	2.2	3.05	23.70

Run	$(\Delta c_W)_{lm}$	$(\Delta c_T)_{lm}$	$K_W a$	$K_T a$	$(H_t)_{OW}$	$(H_t)_{OT}$	m	$\dfrac{m V_W}{V_T}$	$\dfrac{V_T}{m V_W}$
6-3-1	0.008921	0.007178	0.870	1.08	6.66	4.79	1.136	1.27	0.786
2	0.009049	0.007445	0.884	1.07	8.89	4.82	1.112	1.69	0.593
3	0.008604	0.007034	1.32	1.62	4.38	5.52	1.128	0.732	1.37
4	0.008472	0.006975	0.960	1.17	10.4	4.44	1.112	2.15	0.465
5	0.008964	0.007098	0.592	0.747	9.81	4.08	1.150	2.19	0.457
6	0.009361	0.007681	0.603	0.735	13.0	4.15	1.100	2.84	0.353
7	0.009238	0.007751	0.655	0.780	15.3	3.91	1.074	3.53	0.283
8	0.009388	0.008020	0.673	0.787	18.2	3.87	1.049	4.22	0.237
9	0.009242	0.007936	0.717	0.835	21.1	3.65	1.041	5.16	0.194
10	0.009512	0.008225	0.687	0.794	29.7	3.84	1.032	6.90	0.145
11	0.009170	0.006908	1.09	1.45	13.8	3.56	1.192	3.48	0.287
12	0.009019	0.006866	1.79	2.36	8.43	3.80	1.190	2.01	0.497
13	0.008751	0.006643	2.43	3.20	6.22	3.97	1.205	1.43	0.698
14	0.008142	0.006069	3.49	4.68	4.33	4.02	1.231	0.988	1.01
15	0.009096	0.006926	1.78	2.34	13.3	3.82	1.190	3.16	0.317
16	0.009567	0.007346	0.605	0.788	39.2	3.87	1.169	9.08	0.110

brium line is straight is also valid for dilute solutions.

Experimental Procedure

The mass transfer of diethylamine from a continuous water phase to a dispersed toluene phase was investigated. The entering diethylamine concentration in the aqueous phase was held at approximately 1.8 wt. % diethylamine during all the experimental work. No diethylamine was added to the toluene feed; however, on redistillation, a small amount of diethylamine entered the toluene feed and was reported as such. Before operation was started, the liquids in the feed drums were mutually saturated with the opposite solvent. During operation, the two-phase interface was located approximately 1 in. above the top of the packing. Operation was held constant, and flow rates were observed on

the rotameters. After two complete changes of continuous phase, samples of the exit liquids were withdrawn at the sampling taps and were analyzed for diethylamine content. Another set of samples was taken after one more complete change of continuous phase.

In operation, the space in the bell section between the stainless-steel cone section and the bell-section wall (above the packing-support screen) was occupied by air. This space was a dead space and did not in any way affect column operation. In the study of the ¼- and ⅜-in. Raschig rings in the 6-in. I.D. column, a bell section was used constructed of 20-gauge stainless steel identical in design to the Pyrex bell section.

The experiments were carried out at ambient room temperature which varied from 24° to 31° C. as extreme limits. It is evident from the

results of Morello (13) that an over-all variation in temperature of this magnitude is not important for this system in this temperature range. Morello has shown that a variation in temperature of 11.7° C. (from 26.8° to 38.5° C.) results in only an 8% variation in the value of $(H_t)_{OW}$.

Calculation of Mass-transfer Results

A sample table of data for the ½-in. rings in the 6-in. I.D. column is given in Table 1. Values of $K_W a$, $K_T a$, $(H_t)_{OW}$, and $(H_t)_{OT}$, the over-all mass-transfer coefficients and units respectively, were calculated on both the water and the toluene phases. The following relationships were employed for this purpose:

$$K_W a = \frac{N/\theta}{v(\Delta C_W)_{lm}} \qquad (1a)$$

TABLE 2.—HEIGHT OF INDIVIDUAL TRANSFER UNIT AND PACKING PROPERTIES

Nominal packing size, in.	Series	Ratio of column diameter to packing size	Number of packing pieces cu.ft.	Superficial area of packing, a_p (sq.ft.)/(cu.ft.)	Void fraction, F, (cu.ft.)/(cu.ft.)	a_p/F^3, ft.	$(H_t)_W$, ft.	$(H_t)_T$, ft.
¼	3–1	12.0	87,330	214	0.600	992	0.90	2.10
¼	4–1	16.0	87,100	214	0.601	984	0.70	2.10
¼	6–1	24.0	88,370	217	0.595	1029	0.70	2.15
⅜	3–2	8.0	22,660	143	0.602	657	0.90	2.15
⅜	4–2	10.7	24,150	153	0.575	804	0.70	2.10
⅜	6–2	16.0	23,490	149	0.587	735	1.20	2.40
⅜	6–2A	16.0	23,770	150	0.582	763	1.20	2.40
½	3–3	6.0	8,120	95.4	0.653	343	1.20	4.70
½	4–3	8.0	8,800	103	0.624	426	1.00	4.05
½	4–3A	8.0	8,800	103	0.624	426	1.40	4.70
½	6–3	12.0	8,720	103	0.627	416	1.25	3.80
⅝	6–4	9.6	4,880	90.1	0.676	292	1.40	5.20
¾	3–5	4.0	2,460	60.8	0.721	162	1.95	7.20
¾	6–5	8.0	2,800	69.1	0.682	218	1.60	6.10
1	6–6	6.0	1,130	63.0	0.682	198	1.95	8.10
Unpacked	3–7	....	......	.....	.....	...	3.3	14
Unpacked	6–7	....	......	.....	.....	...	3.3	18

Note: Packed volumes: 3-in. I.D. column = 0.202 cu.ft.
4-in. I.D. column = 0.355 cu.ft.
6-in. I.D. column = 0.789 cu.ft.

$$K_T a = \frac{N/\theta}{v(\Delta C_T)_{lm}} \tag{1b}$$

$$(H_t)_{oW} = \frac{V_W}{K_W a} \tag{2a}$$

$$(H_t)_{oT} = \frac{V_T}{K_T a} \tag{2b}$$

where:

N/θ = lb. moles of diethylamine transferred/hr.

v = effective tower volume (cu. ft. of packed bed)

$(\Delta C)_{lm}$ = log mean concentration difference, lb. moles/cu.ft.

V = flow rate, cu.ft./(hr.)(sq.ft.) of empty tower cross section

Subscripts W and T refer to the water and toluene phases respectively

The rate of transfer of diethylamine between the two phases, N/θ, was based on the amount of diethylamine gained by the toluene phase per unit time. Since the direction of mass transfer was from the water to the toluene phase, the change in the diethylamine concentration of the water phase, in passing through the tower, was small. Thus, in the calculation of N/θ based on the water phase, it is necessary to take differences between two relatively large numbers. The loss in significant figures results in poor accuracy in the calculation of N/θ based on the water phase. Inasmuch as the toluene feed was relatively free of diethylamine, a greater concentration change occurred in the toluene phase passing through the tower. Thus, the accuracy of calculating the over-all mass-transfer coefficients is greater if the rate of diethylamine trans-

fer, N/θ, is based upon the amount of diethylamine gained by the toluene phase rather than upon the amount lost by the water phase per unit time.

The approximate log mean concentration difference, $(\Delta C)_{lm}$, was used in place of a more exact relation, the precision of which is not justified in this case. Wiegand (24) has pointed out that the error in using the approximate form is less than 0.1% for a 10% difference and only 1.4% for a 50% difference between the approximate and the true values of the concentration at the interface between the two liquids. Accordingly, $(\Delta C_W)_{lm}$ and $(\Delta C_T)_{lm}$ were calculated by the use of the following equations:

$$(\Delta C_W)_{lm} = \frac{\Delta C_{W1} - \Delta C_{W2}}{\ln \frac{\Delta C_{W1}}{\Delta C_{W2}}} \tag{3a}$$

$$(\Delta C_T)_{lm} = \frac{\Delta C_{T1} - \Delta C_{T2}}{\ln \frac{\Delta C_{T1}}{\Delta C_{T2}}} \tag{3b}$$

where:

$$\Delta C_{W1} = C_{W1} - H_1 C_{T1}$$
$$\Delta C_{W2} = C_{W2} - H_2 C_{T2}$$
$$\Delta C_{T1} = \frac{\Delta C_{W1}}{H_1}$$
$$\Delta C_{T2} = \frac{\Delta C_{W2}}{H_2}$$

H = equilibrium-distribution coefficient

TABLE 3.—PHYSICAL PROPERTIES OF RASCHIG RINGS

Nominal packing size, in.	Outside diameter, in.	Wall thickness, in.	Length, in.	Weight/ piece, g.	Volume/ piece $\times 10^5$, cu.ft.	Surface area/ piece $\times 10^5$, sq.ft.
¼	0.243	0.055	0.244	0.308	0.458	0.245
⅜	0.384	0.080	0.397	1.19	1.76	0.633
½	0.538	0.517	0.517	2.88	4.27	1.18
⅝	0.652	0.099	0.667	4.39	6.64	1.85
¾	0.756	0.129	0.772	7.75	11.3	2.46
1	1.14	0.136	1.14	15.0	28.3	5.59

Subscripts W and T refer to the water and toluene phases, respectively.

Subscripts 1 and 2 refer to the top and bottom of the column, respectively.

Values of H, the equilibrium-distribution coefficient, were taken from a large-scale plot of the equilibrium-distribution data for the toluene-diethylamine-water system (13). Values of m, the slope of the equilibrium curve, were also available from the work of Morello (13).

A material balance was made on the amount of diethylamine entering and leaving the column in the extract and raffinate. The balance was made in the usual manner, and the per cent of error indicated by:

$$\% \text{ error} = \frac{\text{diethylamine in} - \text{diethylamine out}}{\text{diethylamine in}} \quad (4)$$

The over-all height of a transfer unit was resolved into the individual film values for each of the packing sizes used in each of the three columns by the use of the following equations:

$$(H_t)_{OW} = (H_t)_W + (H_t)_T \frac{mV_W}{V_T} \quad (5a)$$

$$(H_t)_{OT} = (H_t)_T + (H_t)_W \frac{V_T}{mV_W} \quad (5b)$$

where m is the slope of the equilibrium curve dC_W/dC_T (C_W referring to the equilibrium solute concentration in phase W and C_T to that in phase T), and V_W and V_T are the flow rates of

phases W and T, respectively.

Equations (5a) and (5b) provide an interesting method for correlating extraction data under widely varying conditions of flow. Thus, if $(H_t)_{OW}$ is plotted against mV_W/V_T, a straight line is obtained, the intercept of which on the $(H_t)_{OW}$ axis will represent the individual film unit, $(H_t)_W$, and the slope of the line will be $(H_t)_T$. In order for the intercept and slope to be true values of the individual film units, the following conditions must be met:

The equilibrium line is straight; i.e., $m = dC_W/dC_T = C_W/C_T$. This condition is generally true in dilute solutions.

The operating line is straight; i.e., the volume of each solvent phase is constant throughout the column.

The individual film transfer units are constant and do not vary with flow rates, column packing, and solute concentrations in the column.
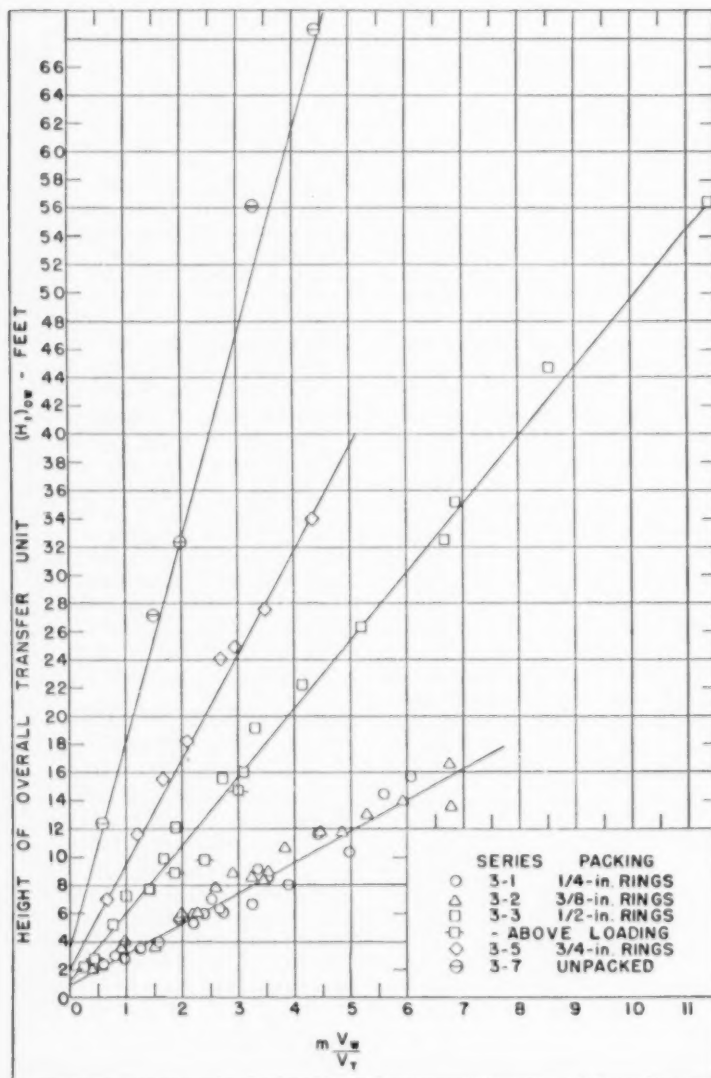
Column Behavior and Packing Size

The appearance and behavior of the column with respect to variation in packing size were similar to those observed by previous investigators (3 and 11). A large number of photographs were taken of the column during operation with the various packing sizes. In most cases, the top of each picture corresponded to a point approximately 1 or 2 in. below the top of the packing. A number of typical photographs are included in Figure 3.

The column operation with ring sizes $\frac{1}{2}$ in. and larger is markedly different from that with $\frac{1}{4}$-in. rings. In the case of the former, the droplets of the dispersed phase appear to be well defined, of nearly spherical shape, and of a relatively constant size distribution independent of packing size. Figure 3a is illustrative of column behavior with ring packings $\frac{1}{2}$ in. and larger.

The analysis of a large number of photographs not included here revealed the following type of behavior for operation with ring sizes $\frac{1}{2}$ in. or larger. At low dispersed-phase flow rates, large portions of the packing (as observed at the wall) appear to be relatively inactive, with few droplets passing through these portions. As the dispersed-phase flow rate is increased, the utilization of the packing by the droplets becomes more uniform. A marked increase of the dispersed phase holdup is also evident. The droplet size of the dispersed toluene phase appears to be independent of increases in the flow rate of the dispersed phase. At a constant continuous-phase flow rate, the droplet size of the dispersed phase is approximately equal for each of the packing sizes $\frac{1}{2}$ in. or larger.

The loading point was not reached for any of the packing sizes larger than

Fig. 4. Mass-transfer results, $(H_t)_{OW}$, in 3-in. I.D. column.

½ in., owing to the limited flow capacity of the equipment. Operation above the loading point was observed for ½-in. rings in the 3-in. I.D. column. The dispersed-phase holdup is greatly increased above the loading point. The drops of the dispersed phase, although bunched together, are distinct and well defined, and no appreciable droplet coalescence occurs. In fact, the droplet size of the dispersed phase in operation above the loading point appears to be slightly smaller than that in operation below the loading point. Figure 3h is illustrative of column behavior above the loading point.

The type of flow observed with the ¼-in. rings differs greatly from that described above. The droplets rise through the packing as irregularly shaped globules. Each droplet hangs in the interstices of the packing until impacted by one from behind. The drops emerge from the top of the packing at only a few different points. The position of these active points on the top of the packing varies with time, and the droplets emerge from these points as streams of well-defined droplets.

Figures 3b and 3c are illustrative of the type of flow encountered for ¼-in. rings. The analysis of a number of photographs revealed that the drop size of the dispersed phase is independent of increases in the dispersed-phase flow rate. The holdup of the dispersed phase is increased as the dispersed-phase flow rate is increased. A study of Figures 3b and 3c reveals that an increase in the continuous-phase flow rate is accompanied by some increase in the droplet size and a definite increase in the holdup of the dispersed phase. Upon a comparison of Figures 3b and 3c for ¼-in. rings with Figures 3a for ½-in. rings, one concludes that the average droplet size in the ¼-in. packing is much larger than that in packing sizes ½ in. or larger.

The type of flow observed with the ⅜-in. rings may be classified as a transition between the droplet behavior through ¼-in. rings and that through packings ½ in. or larger. At low flow rates, the droplet flow through ⅜-in. rings is much the same as observed for packings ½ in. or larger. However, as the dispersed- or the continuous-phase flow rate is increased, the droplet size increases markedly, and the type of flow becomes similar to that characteristic of the ¼-in. rings. This is well illustrated by the contrast between Figures 3d and 3e and between Figures 3f and 3g. It is important to note that at comparable flow rates, the drop size of the dispersed phase in the ⅜-in. rings is somewhat smaller than that in the ¼-in. rings. This is evident from a comparison of Figures 3f and 3b.

In summary, the following conclusions emerge from the preceding photographic analysis:

The droplet size of the dispersed phase is independent of packing size for packings ½ in. or larger.

The droplet size of the dispersed phase is greater for ¼-in. rings than for any of the larger sizes of packing.

The ⅜-in. rings behave as a critical or a transition packing between ¼-in. rings and packings ½ in. or larger.

Results of Mass-transfer Studies

Table 1 lists the experimental results obtained for ½-in. rings in the 6-in. I.D. column.

The experimental results are plotted as $(H_t)_{OW}$ vs. mV_W/V_T in accordance with Equation (5a). The results for each of the three columns are shown in Figures 4, 5, and 6. Plotting the results in this way indicates the validity of a straight-line relationship for a wide range of flow ratios. Positive values of the slope and intercept are obtained in every case.

The experimental results may also be plotted as $(H_t)_{OT}$ vs. V_T/mV_W in accordance with Equation (5b). A straight-line relationship is indicated for a wide range of flow ratios. Positive values of the intercept and slope are obtained in every case.

The relationship between the $(H_t)_{OW}$ and the corresponding $(H_t)_{OT}$ plots, assuming that the individual H_t's are relatively independent of the flow rates, is that the slopes of the lines in Figures
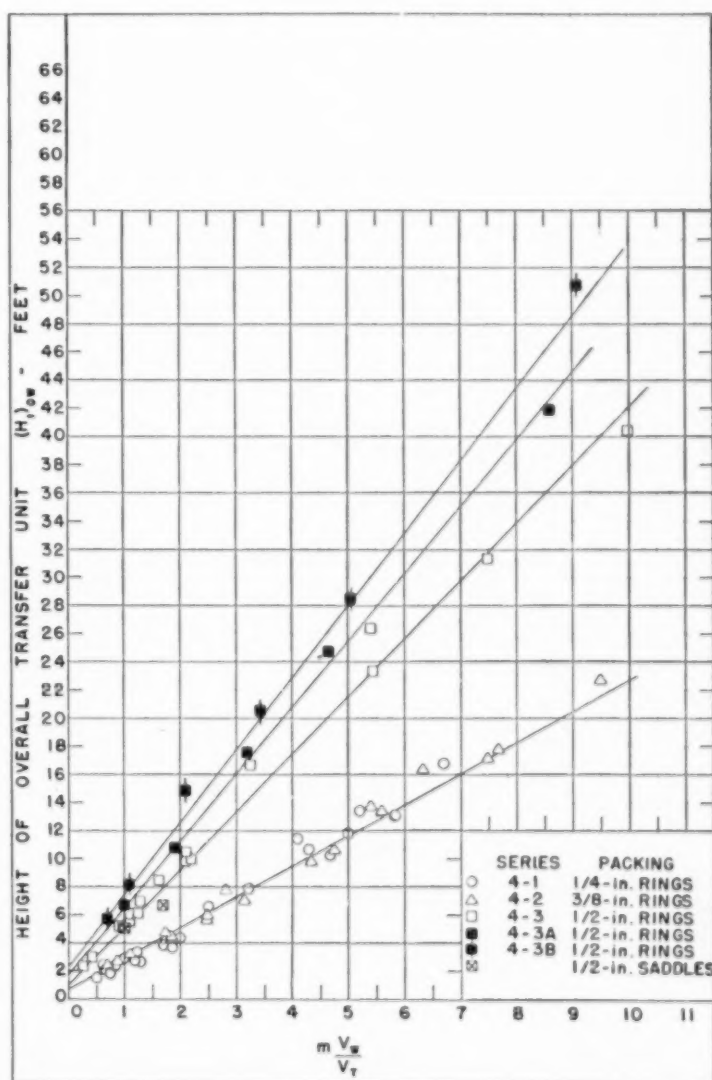

Fig. 5. Mass-transfer results, $(H_t)_{OW}$, in 4-in. I.D. column.

SERIES	PACKING
4-1	¼-in. RINGS
4-2	⅜-in. RINGS
4-3	½-in. RINGS
4-3A	½-in. RINGS
4-3B	½-in. RINGS
	½-in. SADDLES

4, 5, and 6 become the intercepts of the lines in the corresponding $(H_t)_{OT}$ plots and, conversely, that the intercepts in Figures 4, 5, and 6 become the slopes in the corresponding $(H_t)_{OT}$ plots. This condition is verified within the experimental error of the data. The values for the individual H_t's for each of the packings are given in Table 2. These values were obtained from large-scale plots of the results. Straight lines were drawn to fit the points of the least probable error as closely as possible. The value of the individual film-transfer unit reported for a given packing is the average of the values obtained for the slope in Figures 4, 5, and 6 and the intercept in the corresponding $(H_t)_{OT}$ plots and vice versa. In all cases the deviation in the correspondence between the slope

and the intercept value for the individual film H_t is less than 10%.

In the experimental work the loading point was not reached for any of the packing sizes larger than ½ in. The loading velocities for the toluene-water system using ½-in. ring packing may be calculated from the data of Ballard and Piret (1). It is probable that the effect, on the loading velocities, of the change in the physical properties of the toluene-water system due to the presence of diethylamine as a solute, is small. A comparison of the calculated loading velocities with the data for ½-in. rings in the 3-in. I.D. column (series 3-3) reveals that runs 7, 8, 9, and 10 are above the loading point. These runs are shown as solid points in Figure 4. It is evident that lower values of $(H_t)_{OW}$

are obtained for these points.

In discussing droplet behavior with respect to packing size, it was pointed out that for a specific dispersed-phase flow rate, the droplets of the dispersed phase in operation above the loading point are smaller than in operation below the loading point. However, it would seem reasonable that the bunching together of the droplets in operation above the loading point would largely nullify any increases in the interfacial contact area due to smaller drop size. In photographs of column operation above the loading point, there appeared to be a number of very small droplets of the dispersed phase. It is possible that the breakdown of toluene droplets at high flow rates as observed by Row, Koffolt, and Withrow (16) may be responsible for the smaller over-all H_t values obtained for these points. However, the data above the loading point are too meager to warrant any positive statements concerning this behavior.

Reproducibility of Results

The results obtained for a specific packed bed are reproducible within 10% on a relative basis. The internal consistency of the data, within the experimental error of measurement, is well represented by Figures 4 through 6. It is important to note that the data for the column operated unpacked are not so accurate as the remainder of the data. This is a direct consequence of the low efficiency of an unpacked column for mass transfer in the water-diethylamine-toluene system. In the unpacked column, the smaller values which are obtained for the rate of mass transfer are subject to relatively greater errors than in the other modes of operation. This is reflected by a lower accuracy in the calculated H_t values for the unpacked column.

An attempt was made to evaluate the reproducibility of the results upon repacking the packed bed. One series of runs (series 4-3A) was made with ½-in. rings in the 4-in. I.D. column in which the column was repacked. The results are compared with the original series of runs (series 4-3) in Figure 5. A 15% increase in $(H_t)_{OW}$ is indicated. The column-diameter-to-packing-size ratio for these runs is 8.0. It is of interest to note that Dell and Pratt (7) and Gayler and Pratt (8) encountered similar difficulties in the reproducibility of flooding and holdup data respectively, using ½-in. ring packing in a 3-in. I.D. column. In their case the ratio of column diameter to packing size was only 6.0. Gayler and Pratt suggested that similar difficulties are to be expected in the reproducibility of the results of mass-transfer studies upon repacking
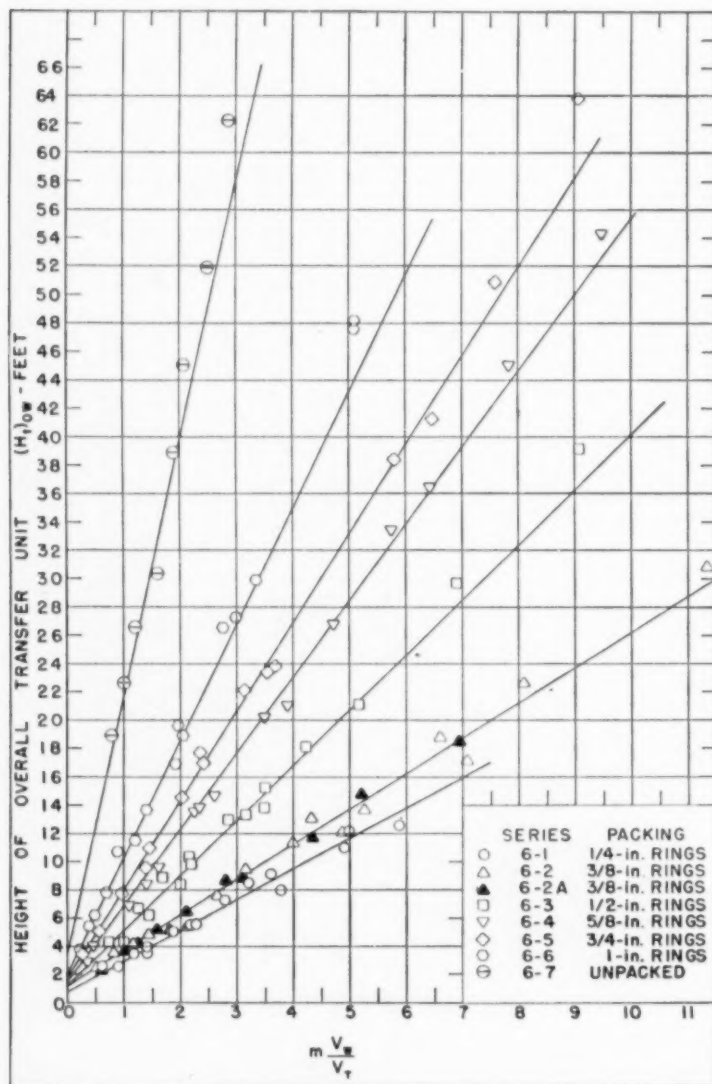
Fig. 6. Mass-transfer results, $(H_t)_{OW}$, in 6-in. I.D. column.

SERIES	PACKING
6-1	1/4-in. RINGS
6-2	3/8-in. RINGS
6-2A	3/8-in. RINGS
6-3	1/2-in. RINGS
6-4	5/8-in. RINGS
6-5	3/4-in. RINGS
6-6	1-in. RINGS
6-7	UNPACKED

the column. These difficulties are probably caused by random differences in the alignment of the pieces of packing. Poor reproducibility of results is probably to be expected for check runs, in which the column has been repacked in the case of column diameter to packing-size ratios of 8.0 or less.

A series of check runs (6-2A) was also carried out for ⅜-in. rings in the 6-in. I.D. column, for which the column was repacked. These results are shown in Figure 6. In this case (a ratio of column diameter to packing size of 16.0) the results were reproducible well within the limits of experimental accuracy with an average deviation of ±3%.

End Effect

The packed height of each of the columns used was held constant at a value of 45.5 in. for all runs with the exception of those in series 4-3B (with ½-in. rings in the 4-in. I.D. column). For series 4-3B, the packed height of the column was 73.0 in. The column was not repacked for series 4-3B. An extra 2 ft. of 4-in. I.D. Pyrex-glass pipe was added to the top of the column at the conclusion of series 4-3A, and the extra ½-in. ring packing necessary to reach a packed height of 73 in. was dumped wet (as described in a preceding section). The mass-transfer results are shown in Figure 5. An increase of 10% in the value of $(H_t)_{OW}$ for series 4-3B is indicated upon comparison with series 4-3A. Thus the presence of a slight over-all positive end effect is indicated. End effects for mass transfer in spray columns have been observed by several investigators (9 and 12). Fur-

ther work is necessary to establish the magnitude and the nature of any possible end effects which may occur in packed columns.

Flow-rate Effects

In order for the slopes and the intercepts of the lines in Figures 4 through 6 to be equal to the values of the respective individual film H_t's, the individual film H_t's must be independent of the flow rates. For the individual film H_t's to be an appreciable function of the flow rates, the lines fitting the experimental data according to Equations (5a) and (5b) must be curved. However, there is no appreciable curvature in the lines of Figures 4 through 6 over the range of flow rates studied. Inasmuch as the experimental data were obtained at reproducible flow rates, it is possible to analyze the data for the effect of the individual flow rates.

Since Ka values are very sensitive to fluctuation in flow compared to H_t values, the Ka values will be used in analyzing the experimental data for the effect of flow rates. A typical plot of K_Wa vs. V_T with V_W as a parameter using logarithmic coordinates is shown in Figure 7, where K_Wa is plotted vs. V_T for ⅝-in. rings in the 6-in. I.D. column. Figure 7 is typical of the data for all packing sizes ½ in. or larger. The average slope for these plots is slightly less than unity, an indication that within the accuracy of the data, K_Wa increases linearly with an increase in the dispersed-phase flow rate, V_T. However, an interaction effect of the flow rates is also indicated, inasmuch as the slopes of the plots (for all the

packing sizes ½ in. or larger) decrease as V_W is decreased. It is evident from Figure 7 that K_Wa increases only slightly for an increase in the continuous-phase flow rate, V_W, in contrast to the effect of increasing the dispersed-phase flow rate.

Similar plots of K_Wa vs. V_T, using logarithmic coordinates, were prepared for ¼-in. rings in each of the columns used. The average slope for all the plots for ¼-in. rings is approximately 0.8. This indicates that the over-all mass-transfer coefficient, K_Wa, is a function of $V_T^{0.8}$. Some interaction effect of the flow rates is also indicated, inasmuch as the slopes of the plots decrease as V_W is decreased.

It might appear that the following equations:

$$(H_t)_{OD} = C_1 + C_2 \left(\frac{V_C}{V_D} \right)^n \left(\frac{V_D}{mV_C} \right)$$

$$(6a)$$

$$(H_t)_{OC} = C_2 \left(\frac{V_C}{V_D} \right)^n + C_1 \left(\frac{mV_C}{V_D} \right)$$

$$(6b)$$

are applicable to the data for ¼-in. rings where C_1 and C_2 are constants and n must be determined experimentally. However, the use of these equations for the toluene-water system requires the assumptions that the dispersed-phase individual film-transfer unit, $(H_t)_T$, is a constant independent of flow rates and that the entire dependence upon the flow rates takes the form of a power function of V_W/V_T in the case of the individual film-transfer unit of the continuous phase, $(H_t)_W$. The data of Colburn and Welsh (5), upon which Equations (6a) and (6b) are based, scatter widely. The data of Laddha and Smith (10) show some variation of the dispersed-phase individual film-transfer unit with flow rate. Thus, the validity of Equations (6a) and (6b) for separating values of the over-all H_t's into individual film H_t's is open to question. At any rate, for the toluene-diethylamine-water system, the value of $(H_t)_W$ is so small compared to the value of $(H_t)_T$ that the use of Equations (5a) and (5b) to correlate the data does not result in any appreciable error.

Similar plots of K_Wa vs. V_T, using logarithmic coordinates, corroborate the transition behavior of ⅜-in. rings. At low toluene flow rates, the behavior is similar to that of packing sizes ½ in. and larger, and at corresponding flow rates the values for K_Wa are smaller than those for ¼-in. rings. At higher toluene flow rates, the values of K_Wa for ⅜-in. rings become equal to those for ¼-in. rings. Some interaction effect of the
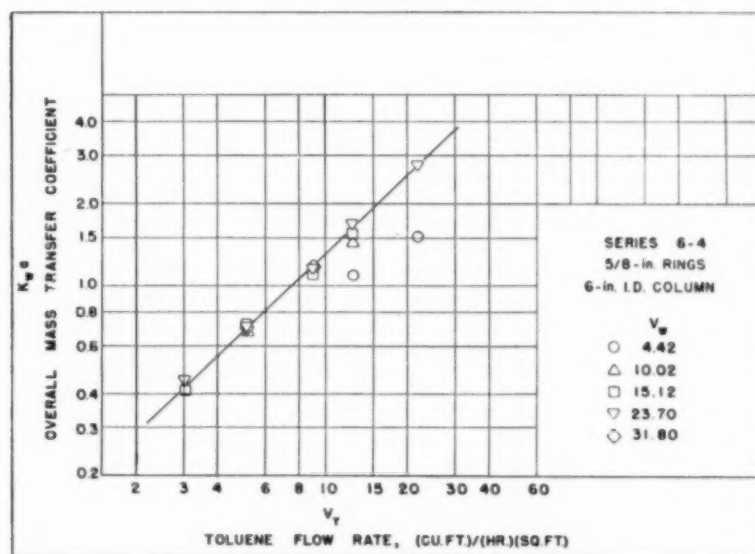

Fig. 7. Effect of dispersed-phase flow rate on the mass-transfer coefficient, (K_Wa), for ⅝-in. rings.

flow rates is also indicated for ⅜-in. rings.

The foregoing results suggest that the effect on the mass-transfer coefficients of variations in flow rates is complex for packed columns. Nevertheless, the use of Equations (5a) and (5b) to correlate the data is within experimental error in this investigation.

Effect of Packing Size on Mass Transfer

The effect of packing size on mass transfer is well illustrated in Figures 4 through 6. Packing sizes ½ in. or larger possess markedly higher H_t values than do the smaller packings. Figures 4 through 6 indicate a definite trend for packings ½ in. or larger in that for a specific column diameter, the value of H_t increases as the size of the packing is increased. The constancy of the droplet size of the dispersed phase for packings ½ in. or larger is discussed in connection with the photographic analysis of column behavior. For these packing sizes, the droplets of the dispersed phase are smaller than the void spaces of the packings. For a constant drop size of the dispersed phase the holdup of the dispersed phase governs the mean velocity of droplet rise through the packing and thereby fixes the time of contact (and the interfacial contact area) of the droplets with the continuous phase. Gayler and Pratt (8) have shown that the holdup of the dispersed phase decreases as the size of the packing is increased. This explains why less efficient mass transfer is obtained with an increase in packing size for packings ½ in. or larger. This behavior supports the concept that one of the main functions of packings ½ in. or larger is to act as a series of baffles in the column. Furthermore, for a specific system and a specific type of packing, one would expect the individual film H_t's to be correlatable with some factor characterizing the size of the packing. A correlation of this type for the toluene-diethylamine-water system with porcelain-ring packings is discussed in a subsequent section of the paper.

In order to explain the greater mass-transfer efficiency of the ¼-in. packing, one must consider the over-all effect of several important factors. Gayler and Pratt (8) have shown that the holdup of the dispersed phase for ¼-in. ring packing is greater than that for any of the larger packing sizes. This increase in the holdup of the dispersed phase for ¼-in. rings results in a longer time of contact between the droplets and the continuous phase. On the other hand, the droplet size of the dispersed phase is greater for ¼-in. rings than for any of the other packings. Thus the resultant interfacial contact area is deter-

mined by a balance between two opposing factors. Furthermore, the deformation of the droplets and the collisions between successive droplets in passing through the small void spaces of the ¼-in. rings probably results in an increase of the internal circulation within the drops. From the work of Sherwood et al. (19) and Licht and Conway (12) with single drops in spray columns, this behavior may be expected to effect an increase in the mass-transfer coefficient, K, and thus decrease the value of H_t. The over-all result of these factors is a greater efficiency of mass transfer for ¼-in. rings than for any of the larger sizes of packing.

Figures 4 through 6 show that there is little difference between the values of H_t for ⅜- and ¼-in. rings. In explaining this close relationship, one must reconsider the same factors which influence the mass-transfer efficiency of the ¼-in. rings. The holdup of the dispersed phase for ⅜-in. rings, being slightly less than that for ¼-in. rings, effects a decrease (compared to ¼-in. rings) in the time of contact for the droplets of the dispersed phase as compared with those of the continuous phase. On the other hand, the droplet size of the dispersed phase is somewhat smaller for ⅜-in. rings than for the ¼-in. ring packing. Thus, the over-all effect of these two opposing factors is approximately the same as that for ¼-in. rings. Furthermore, on the basis of the work of Sherwood et al. (19), one may expect less internal circulation within the droplets for ⅜-in. rings, inasmuch as these droplets are smaller than those for ¼-in. rings. The preceding analysis qualitatively justifies the mass-transfer results for these two packings.

The least efficient mode of operation for mass transfer in the toluene-diethylamine-water system is the operation of an unpacked column over the same range of flow rates as is used for the packed columns. The series of runs with the columns operated unpacked show how much effect the presence of the most inefficient packing size (1-in. rings) has on the rate of mass transfer compared to an unpacked column. Using the unpacked column for this system, one would require a column height approximately twice as great as the packed height of 1-in. rings necessary to secure a specific rate of mass transfer.

The only known data for the toluene-diethylamine-water system determined with another industrial packing consists of two experimental runs reported by Wehn and Franke (23), who used a 4-in. I.D. column with a packed height of 5 ft. of ½-in. Berl saddles. These data are not directly comparable to the present investigation because the inlet

solute concentration of the water feed was approximately twice that used in the present investigation. Nevertheless, the relative magnitudes of the H_t values reported for the ½-in. Berl saddles are comparable with the data obtained for ½-in. rings in the 4-in. I.D. column. The data of Wehn and Franke are shown on Figure 5.

Effect of Column Diameter on Mass Transfer

The results of the present investigation indicate that a ratio of column diameter to packing size greater than 8.0 should be maintained in the design of packed liquid-liquid extraction columns. For a specific packing size, the effect of operating with a column-diameter-to-packing-size ratio equal to or less than 8.0 is an increase in over-all H_t values, i.e., a decrease in mass-transfer efficiency. This is due to a "wall effect," which results in greater than normal values for the void fraction of the packing. In Table 2, the values of the void fractions for ⅜-, ½-, and ¾-in. rings in the 3-in. I.D. column are larger than those for the corresponding packings in the 6-in. column. Furthermore, during the operation of the column for column-diameter-to-packing-size ratios of 6.0 or less, a considerable wall effect is visible. Some channeling takes place, with most of the flow of the dispersed phase occurring near the walls of the column. The effect of an increase in the void fraction of the packing is to decrease the holdup of the dispersed phase in the packing. The result is a decrease in the time of contact (and the interfacial contact area) for the droplets of the dispersed phase with the continuous phase. As a consequence, the mass-transfer efficiency of the column is decreased. This effect is shown in Figure 8 for ½- and ¾-in. rings. As discussed in a preceding paragraph, wall effect also results in a serious lack of reproducibility in flooding (7), hold-up (8), and mass-transfer results upon repacking the column. For these reasons, in future work with packed liquid-liquid extraction columns, the ratio of the column diameter to the packing size should be maintained at a value greater than 8.0.

A consideration of the H_t values for ⅜-in. rings in the 3-, 4-, and 6-in. I.D. columns reveals that the individual film H_t's vary with column diameter. In Table 2 smaller values of the individual film H_t's are shown for ⅜-in. rings in the 4-in. I.D. column than in the 3-in. I.D. column. This is probably due to wall effect, as discussed in the preceding paragraph. However, the values for the individual H_t's for ⅜-in. rings in the 6-in. I.D. column are greater than those
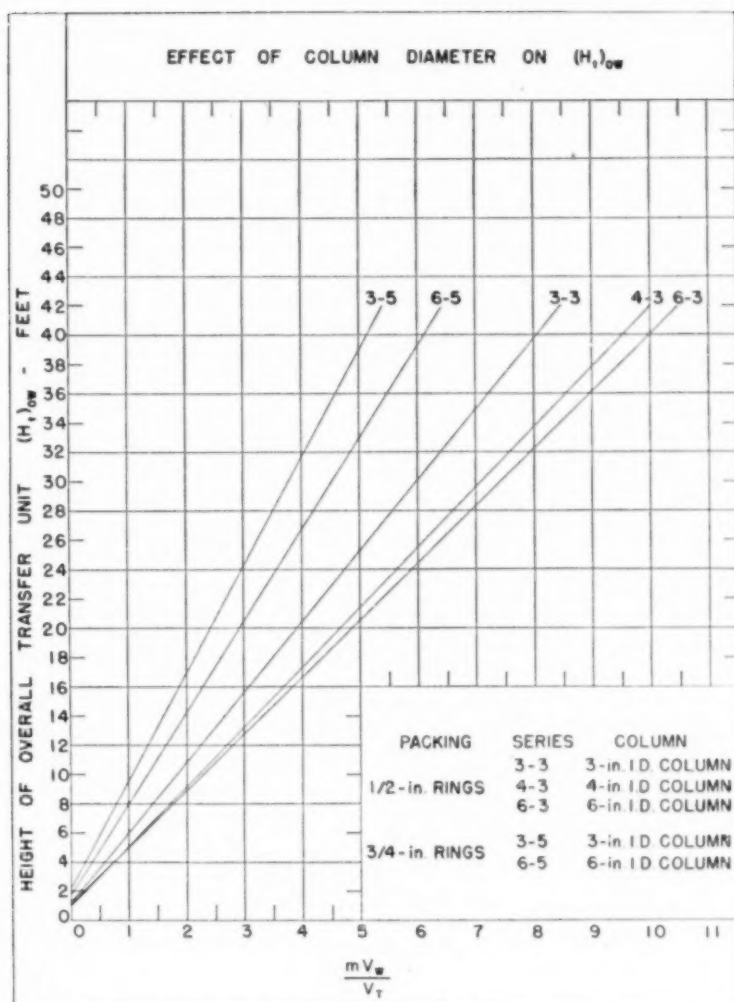
Fig. 8. Effect of tower diameter on mass transfer, $(H_t)_{ow}$.
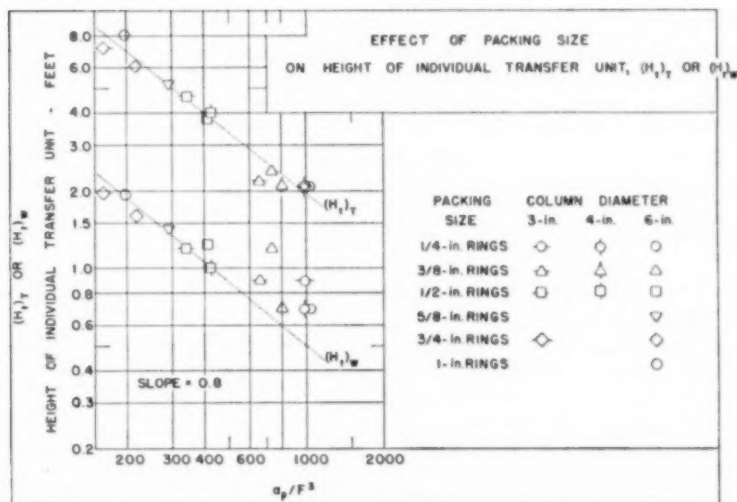

Fig. 9. The effect of packing size and tower diameter on mass transfer, $(H_t)_T$ or $(H_t)_w$.

for either of the other columns. This may be in some way connected with the transition behavior of $\frac{3}{8}$-in. ring packing.

The effect of column diameter on the results for the unpacked column is interesting. In the 3-in. I.D. column, a value approximately 20% smaller for the individual film $(H_t)_T$ is obtained compared with that for the 6-in. I.D. column. Photographs of column operation for the unpacked columns showed a flattened, elliptical shape for the droplets of the dispersed phase. This drop shape implies that the drag on the top half of the drops may be sufficient to set up some local recirculation of the continuous phase in both columns. As discussed by Morello and Poffenberger (15), one might expect this effect to increase with an increase in column diameter for spray columns.

Correlation of Effect of Packing Size on Mass Transfer

For a specific system and a specific type of packing, the feasibility of correlating the individual film H_t's with some factor characterizing the size of the packing (in this case, a_p/F^3) is discussed above. A correlation of this type for the toluene-diethylamine-water system with porcelain-ring packings is presented in Figure 9 for packings $\frac{1}{2}$ in. or larger. In view of the markedly different type of flow for $\frac{1}{4}$- and $\frac{3}{8}$-in. rings, the scatter shown by the points for these packings is to be expected. In industrial liquid-liquid–extraction packed-column installations, the likelihood of packings smaller than $\frac{1}{2}$-in. rings ever being used is small.

For the toluene-diethylamine-water system, the resistance of the toluene film to mass transfer in packed columns is much greater than that of the water film. The correlation presented here is applicable only to the system studied in this investigation. Further work with other systems is necessary before a more general correlation can be developed.

Summary

The mass transfer of diethylamine from a continuous water phase to a dispersed toluene phase was investigated. The over-all height of a transfer unit was resolved into the individual film values for each of the packing sizes used in each of the three columns. For packing sizes $\frac{1}{2}$ in. or larger, a definite trend was indicated in that, for a specific column diameter, the value of H_t increased as the size of the packing was increased. The values of H_t obtained for the $\frac{1}{4}$- and $\frac{3}{8}$-in. rings were smaller than those for any of the other packings.

Photographs of the dispersed-phase droplet behavior in the packing during

column operation revealed that the droplet size of the dispersed phase is (a) independent of packing type and size for packings $\frac{1}{2}$ in. or larger and (b) greater for $\frac{1}{4}$-in. rings than for any of the larger sizes of packing. The $\frac{3}{8}$-in. rings behaved as a critical or a transition packing between $\frac{1}{4}$-in. rings and the sizes of packing $\frac{1}{2}$ in. or larger.

The photographic analysis permitted an explanation of the mass-transfer results in terms of changes in the interfacial contact area due to the effect of the packing size on the dispersed-phase droplet size and on the holdup of the dispersed phase. A correlation of the effect of packing size on the individual film height of a transfer unit was developed for packing sizes $\frac{1}{2}$ in. or larger.

The mass-transfer results indicated that a ratio of column diameter to packing size of greater than 8.0 should be maintained in the design of packed countercurrent liquid-liquid extraction columns. For a specific packing size, the effect of operating with a column-diameter-to-packing-size ratio equal to or less than 8.0 is an increase in the over-all H_t values, i.e., a decrease in the mass-transfer efficiency.

Acknowledgment

The authors wish to acknowledge the suggestions and advice of Dr. V. S. Morello of the Dow Chemical Company. The help of Mr. Kurt Li in the development of the photographic technique and Mr. Leroy Smith in the machine work is sincerely appreciated.

Notation

a = interfacial contact area/cu.ft. of column volume, sq.ft./cu.ft.

a_p = packing surface area/cu.ft. of column volume, sq.ft./cu.ft.

c = solute concentration in bulk stream, lb. moles/cu.ft.

c^* = solute concentration in one phase which would be in equilibrium with observed concentration in the other phase, lb. moles/cu.ft.

$(\Delta C)_{lm}$ = log mean concentration difference, lb. moles/cu.ft.

F = void fraction of the packing

H = equilibrium distribution coefficient, ratio of solute concentration in aqueous phase to that in non-aqueous phase at equilibrium

H_t = height of a transfer unit, ft.

$(H_t)_o$ = height of an over-all transfer unit, ft.

K_T = over-all mass-transfer coefficient based on the equivalent driving force in the toluene phase, lb. moles/(hr.)(sq.ft.)(lb. moles/cu.ft.)

K_{Ta} = over-all mass-transfer coefficient based on the equivalent driving force in the toluene phase, lb. moles/(hr.)(cu.ft.)(lb. moles/cu.ft.)

k_T = toluene film mass-transfer coefficient, lb. moles/(hr.)(sq.ft.)(lb. moles/cu.ft.)

k_{Ta} = toluene film mass-transfer coefficient, lb. moles/(hr.)(cu.ft.)(lb. moles/cu.ft.)

K_W = over-all mass-transfer coefficient based on the equivalent driving force in the water phase, lb. moles/(hr.)(sq.ft.)(lb. moles/cu.ft.)

K_{Wa} = over-all mass-transfer coefficient based on the equivalent driving force in the water phase, lb. moles/(hr.)(cu.ft.)(lb. moles/cu.ft.)

k_W = water film mass-transfer coefficient, lb. moles/(hr.)(sq.ft.)(lb. moles/cu.ft.)

k_{Wa} = water film mass-transfer coefficient, lb. moles/(hr.)(cu.ft.)(lb. moles/cu.ft.)

m = slope of the equilibrium curve, dc_W/dc_T

N = number of lb. moles of solute transferred

T = temperature, ° C.

V = superficial liquid velocity, cu.ft./(hr.)(sq.ft. of column cross section)

v = effective volume of column, cu.ft.

θ = time, hr.

SUBSCRIPTS

C = continuous phase

D = discontinuous phase

lm = logarithmic mean

T = toluene phase

W = water phase

1 = top of column

2 = bottom of column

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Discussion

Leo Garwin (Kerr-McGee Oil Industries, Inc., Oklahoma City, Okla.): I was interested in your finding that the droplet size with the $\frac{1}{4}$-in. rings was larger than that with the $\frac{3}{8}$-in. rings, and I am wondering if it might be due to the selection of the particular nozzle size which you employed. Do you think you might have gotten the regular progression had you used a smaller nozzle diameter?

Irving Leibson: The nozzle diameter employed in this study was approximately 1/10-in. I.D. I do not believe that varying nozzle diameter would have had any effect for this fine packing inasmuch as packing size tended to control the droplet size for this particular packing. However, I do believe that for packing sizes $\frac{1}{2}$ in. or greater one should be extremely careful in choosing the inlet nozzle size for the dispersed-phase distributor. It may be possible that some other optimum nozzle diameter would exist. However, in previous work done by several investigators using the methyl-isobutyl-ketone-water system with acetic acid as a solute, the optimum nozzle diameter was found to be 1/10-in. I.D. Since we had no better information in designing our column, we used that size. But I do not believe that nozzle diameter is an important variable for packing sizes smaller than $\frac{3}{8}$ in.

Presented at A.I.Ch.E. Chicago meeting.

Factors Influencing the Properties

of Spray-Dried Materials

J. A. Duffie and W. R. Marshall, Jr.

University of Wisconsin, Madison, Wisconsin

Part I

The original objective of the research reported in this paper was to extend the study of evaporation from drops as reported by Ranz and Marshall (10) from conditions of a suspended single drop evaporating in a constant-temperature gas stream to conditions which simulate more closely the conditions in a spray dryer, i.e., conditions of a free drop in an atmosphere of changing temperature. However, because certain experimental difficulties prevented completion of all objectives, this paper is a report of a less quantitative study of the effects of certain drying conditions on the product properties of the materials studied. Consequently, it is presented as an additional phase in the continuing study of droplet evaporation as related to spray drying.

The results of the study are reported from two standpoints. One is based on a study of the bulk densities of the materials studied and the manner in which they varied with drying-air temperature, liquid-feed temperature, and feed concentration. The second is based on a classification of materials according to the nature of their spray-dried products.

Few quantitative data are available relating spray-dried-product characteristics to operating variables or to the physical or chemical properties of the materials being dried. Wallman and Blyth (19) reported a study of the effect of operating variables on the product characteristics of sodium silicate but did not extend their findings to other materials. The data of this study are also specific, in a sense, since they apply only to the materials dried and to the special conditions of drying used. However, the trends are considered significant, while the data emphasize the importance of certain factors, particularly the film-forming tendencies of the material, in imparting properties of hollowness and lightness to the spray-dried particles.

A vertical, cylindrical concurrent-flow spray dryer, 8 in. in diam. and 20 ft. high, was developed for experimental studies on the drying of a stream of drops of controlled size. A viscous-jet atomizer was used to produce the drops, and its breakup characteristics in this application were studied. High-speed motion pictures of the mechanism of jet breakup were obtained.

The bulk densities of several materials were studied as a function of air and feed temperatures, feed concentration, and material properties. The bulk densities of a water-dispersible organic dyestuff and sodium silicate were found to decrease with increasing drying-air temperature, with the effect due largely to increased dried-particle size. The bulk densities of the dyestuff, sodium silicate, and sodium chloride were found to decrease slightly with increasing liquid-feed temperature, with the effect due largely to changes in atomization. The bulk density of the dyestuff was found to decrease with increasing feed concentration, with the effect ascribed to an increase in dried-particle size. However, it was concluded that for other materials the effects of feed temperature and concentration can also be opposite to those observed.

Thirteen materials were spray dried, including inorganic salts, the dyestuff, a dispersing agent, whole milk, gelatin, coffee extract, and corn syrup. Photomicrographs of some of the products are shown.

W. R. Marshall, Jr., recently named associate dean of the College of Engineering, University of Wisconsin, and associate director of the Wisconsin Engineering Experiment Station, was employed by Du Pont Co. at its experimental station in Wilmington, Del., from 1941 to 1947. He then returned to the University of Wisconsin, from which he had received a Ph.D. in 1941, as associate professor in chemical engineering. A consultant to industry, his principal field of research is drying and related topics. At the 1950 meeting of A.I.Ch.E., he and J. A. Gerster, of the University of Delaware, received the awards for the best presentation of a technical paper. In December, 1952, Dr. Marshall presented the fourth Annual Lecture of A.I.Ch.E. The subject was Spray Drying and Atomization.

J. A. Duffie is a scientific liaison officer at the Chicago branch office of the Office of Naval Research. He received a B.Ch.E. and an M.Ch.E. from Rensselaer Polytechnic Institute, Troy, N. Y., where he was an instructor of chemical engineering for three years. He was awarded his Ph.D. from the University of Wisconsin in 1951. Before being recalled to naval service, Dr. Duffie was employed as a research engineer at Du Pont's electrochemicals department.
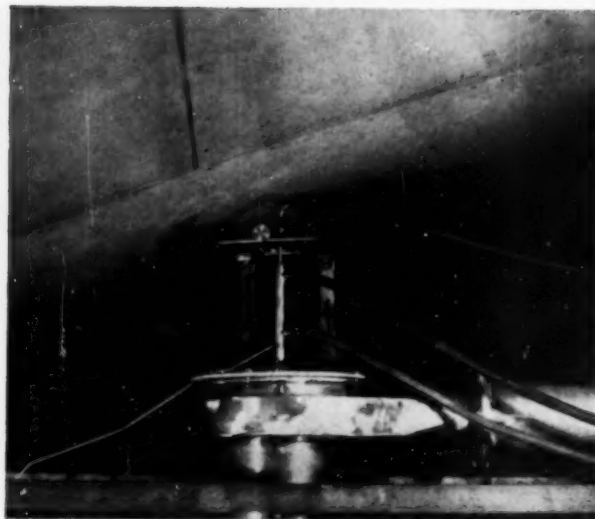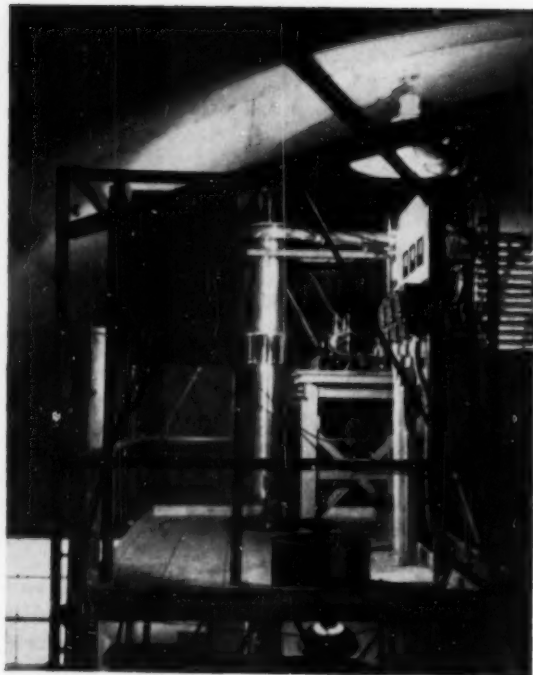
Fig. 1a. Left, General view of experimental drying tower before insulation.

Fig. 1b. Above, View of top of tower and jet atomizer in position.

The experimental dryer used in this study involved the development of a so-called viscous-jet atomizer and a study of its performance in this application. Two distinct experimental phases are considered, the first dealing with a study of the method of drop formation used, and the second with the results of drying experiments.

Description of Experimental Equipment

The somewhat novel design of the dryer used in these experiments was based on the desire to study the drying of drops in free fall and to cover the entire drying range, i.e., the constant- and falling-rate periods. The experimental dryer was designed, therefore, with the following requirements in mind:

1. The path of travel of the drops through the dryer should be well defined.

2. The residence time of the drops in the drying zone should be nearly independent of particle size, or at least measurable.

3. The drops should be as uniform in size as possible.

4. The drops during drying should be accessible for sampling.

5. The air requirements and dryer size should not be excessive.

To meet these requirements as far as possible, a stainless-steel, small-diameter vertical cylindrical drying tower was constructed. A viscous-jet atomizer was used to produce a vertical stream of drops down the dryer concurrent with
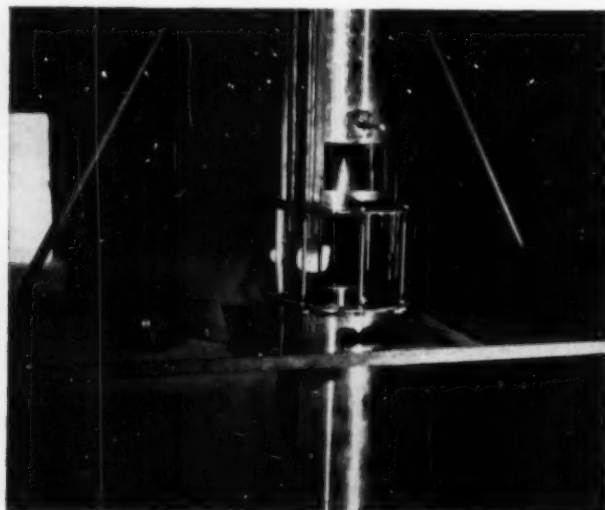
Fig. 1c. Left, View of glass section and sampling port.
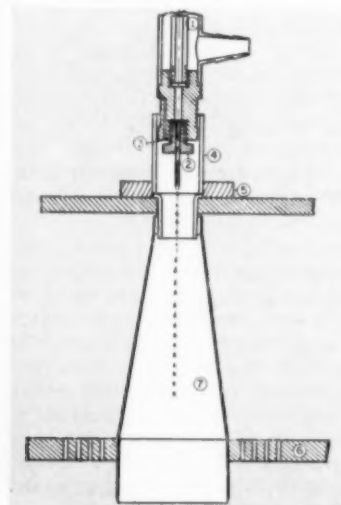
Fig. 2. Section view of top of dryer showing drop entry equipment.
1. jacketed section.
2. nozzle plug.
3. feed thermocouple.
4. glass tubing shield.
5. transite retaining ring and tower top.
6. perforated distributor plate.
7. conical baffle on plate.
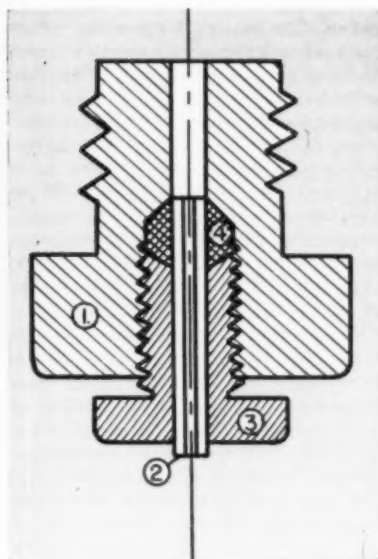
Fig. 3. Assembly of glass capillary nozzle.

1. brass plug.
2. glass capillary tube.
3. packing nut.
4. rubber packing.

the drying air. The purpose of this type of atomizer was twofold: (1) it produced drops without dispersing them into a cone, and (2) the drop size was considered more uniform and better controlled. Photographs of the dryer are shown in Figures 1a, b, c. Details of the top of the tower and the drop-entry section are shown in Figure 2.

The dryer proper was 8 in. I.D. and 19.5 ft. long, composed of alternate lengths of 2.5-ft. stainless-steel duct and of 0.5-ft. Pyrex glass pipe. Auxiliary equipment included electric air heaters, blower, thermocouples, and a temperature recorder for measuring air and feed temperatures.

The jet atomizer consisted of a small-bore capillary tube through which the feed liquid was pumped. The liquid issued as a viscous jet, the jet breaking up into drops under the influence of surface tension. The resulting drops all moved with nearly the same velocity and were relatively uniform in size. The capacity of the nozzle was low. The major disadvantage of this method of drop production was that the drop diameters produced were of the order of twice the diameter of the capillary. This made the production of drops smaller than about 200μ difficult. Capillary sizes down to 200μ were cut from stainless-steel hypodermic tubing, while smaller sizes were drawn from glass capillary tubes. The method of assembly of the latter type is shown in Figure 3, and a photomicrograph of the tip of a nozzle is shown in Figure 4. Glass capillaries with bores as small as 80μ were operated successfully, although some difficulty was encountered owing to clogging or drying of solution in the capillary.

A low-capacity feed system was devised in which a metering unit pumped S.A.E. oil against a mercury-seal column, which in turn forced the feed liquid through the capillary. The feed was filtered through a screen having 40-μ openings, while the feed line was jacketed for temperature control, as shown in Figure 2.

In operation, air from an aircraft-engine supercharger blower was heated in an electric heater and passed into the tower through an annular ring near the top and then through a perforated plate for distribution over the tower cross section. The liquid jet issued from the capillary just above the top of the tower and broke up into drops prior to coming in contact with the hot air. This is shown diagrammatically in Figure 2. The jet atomizer was readily accessible for removal and could be operated outside the dryer when photographic studies of its performance were made.

Theoretical Considerations

The original objectives and the experimental techniques used in this study suggested the need of certain theoretical considerations relating to jet breakup and to the evaporation rate of falling drops:

BREAKUP OF VISCOUS JETS

Lord Rayleigh's pioneering paper (11) formulated mathematically the phenomenon of the breakup of a viscous jet of the type used in this study. Rayleigh stated that jet instability could be attributed to two causes: (1) surface-tension forces for liquid jets issuing into air and (2) dynamical or viscous forces for liquid jets issuing at high velocities into a gas atmosphere or for liquid jets issuing into a second liquid. Only the first cause is of importance in this study, since the jet velocity in all cases was low.

Rayleigh described a liquid jet falling in air as being in a condition of unstable equilibrium and postulated that, in order for this unstable equilibrium to be destroyed, it must be subjected to certain disturbances or displacements, without which the jet would not break up. These disturbances or displacements can be interpreted as being caused by vibrations absorbed by the jet. Rayleigh's analysis assumed that the amplitude of these vibrations increased with time until it exceeded the jet diameter, when breakup would occur. It was further postulated that the shortest breakup time occurred for a disturbance whose wave length corresponded to the condition of greatest instability. This condition was assumed by Rayleigh to occur when the ratio of the surface energy to the jet kinetic energy became a maximum.

A prediction of the principal drop size resulting from jet breakup as a consequence of Rayleigh's analysis is that

$$D_p = 1.9D_j \qquad (1)$$
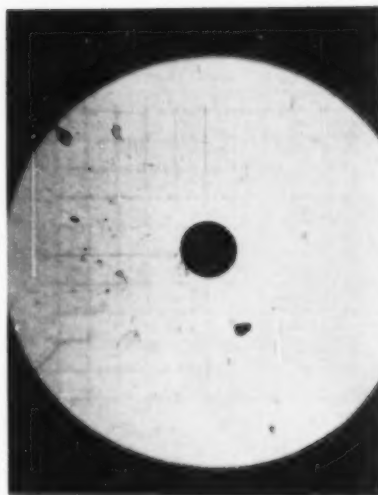
Fig. 4. Photomicrograph of capillary nozzle. Actual diameter of hole is 100μ.

where D_p = drop diameter and D_j = jet diameter. The formation of small satellite drops between larger drops is typical of jet disintegration by this method, although the size and occurrence of these smaller drops can be varied by the impression of controlled vibrations. The use of vibrations was only cursorily considered in this study.

Jet instability due to the action of the surrounding medium did not apply to the slow-moving jets used in this investigation.

A number of authors have reported experimental and analytical studies of jet breakup (3, 7, 13, 15-18), but few have presented data on drop-size distribution to be expected from such atomizers (9). The approach has been a philosophical one, aimed chiefly at aiding the understanding of the mechanism of atomization in other types of atomizers. Few direct applications of the Rayleigh breakup mechanism have been reported in spray-drying or cooling operations.

EVAPORATION LIFETIMES OF DROPS

Sherwood and Williams (14), Ranz and Marshall (10), and Marshall and Seltzer (8) treated the problem of the evaporation lifetimes of pure liquid drops in still air and falling at terminal velocities in the Stokes' law region. Sjenitzer (14a) considered the effect on evaporation of deceleration of drops from a high initial velocity to terminal velocity. It was shown in all instances that corrections to the time of evaporation in still air for laminar motion through the air were negligible for drop sizes below about 100μ. An extension of these considerations was made in this study to cover the range of Reynolds
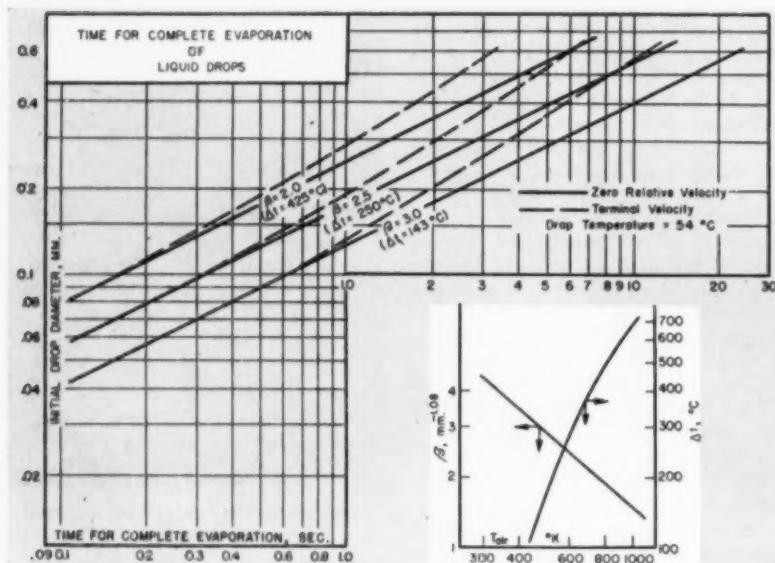
Fig. 5. Evaporation times for pure liquid drop calculated from Equations (3) and (9).

numbers in the intermediate range, and although the results were not actually used in this study, they are reported here for the benefit of other workers in the field.

A heat balance on an evaporating drop of diameter D_p gives for the time of evaporation

$$\theta = \frac{\rho_L \lambda}{2\Delta t} \int_{D_{p2}}^{D_p} \frac{dD_p}{h} . \qquad (2)$$

At $N_{Re} = 0$, the time for complete evaporation is

$$\theta = \frac{\rho_L \lambda D_p^2}{8k_f \Delta t} . \qquad (3)$$

Ranz and Marshall's correlation (10) for heat transfer to drops can be substituted in (2) to give

$$\theta = \frac{\rho_L \lambda}{4k_f \Delta t} \int_{D_{p2}}^{D_{p1}} \frac{D_p dD_p}{[1 + 0.3(N_{Re})^{\frac{1}{2}}(N_{Pr})^{\frac{1}{3}}]} . \qquad (4)$$

Perry gives for the terminal velocity in the transition or intermediate region the empirical equation

$$u_t = \frac{0.153 D_p^{1.14} g^{0.71} (\rho_L - \rho_a)^{0.71}}{\rho_a^{0.29} \mu^{0.43}} . \qquad (5)$$

Substitution of (5) in (4) gives

$$\theta = \frac{\rho_L \lambda}{4k_f \Delta t} \int_{D_{p1}}^{D_{p2}} \frac{D_p dD_p}{1 + \beta D_p^{1.07}} \qquad (6)$$

where

$$\beta = 0.3 \left[0.153 \left(\frac{g\rho_a(\rho_L - \rho_a)}{\mu^2} \right)^{0.71} \right]^{\frac{1}{2}}$$

$$\left[\frac{C\mu}{k} \right]^{\frac{1}{3}} \qquad (7)$$

and from (6)

$$\theta = \frac{\rho_L \lambda}{8k_f \Delta t} \left[(D_{p1}^2 - D_{p2}^2) - 2\beta \int_{D_{p2}}^{D_{p1}} \frac{D_p^{2.07} dD_p}{(1 + \beta D_p^{1.07})} \right] . \qquad (8)$$

In order to put Equation (8) in a convenient form, one can integrate it from an appropriate lower limit, below which the correction is negligible, i.e., 100μ, to D_{p1}. Adding the time for evaporation from 100 to zero μ, as given by (3), gives

$$\theta = \frac{\rho_L \lambda D_p^2}{8k_f \Delta t} \left[1 - \frac{2\beta}{D_{p1}^2} \int_{100}^{D_{p1}} \frac{D_p^{2.07} dD_p}{(1 + \beta D_p^{1.07})} \right] \qquad (9)$$

In Equation (9) the first term is the evaporation lifetime at zero relative velocity, and the second is the correction to account for the terminal velocity when the drop is larger than 100μ.

Here it is convenient to use units of seconds for θ, millimeters for D_p and grams per cubic centimeter for ρ_L. Then, assuming $N_{Pr} = 0.73$, $\rho = 1.00$, $\rho_a = 29(273)/24,000\, T_a$, and $\rho_a << \rho_L$,

$$\beta = 0.071 \left[\frac{\rho_L}{\mu^2 T_a} \right]^{0.36} , \text{mm.}^{-1.07} \quad (10)$$

For a drop temperature of 130° F., Equation (9) was evaluated for three values of the parameter β and the results are plotted in Figure 5. Curves of β vs. air temperature and Δt are also

shown. The spacing between the curves for various Δt values is nearly a linear function of ρ_L. For liquids other than water, evaporation times can be estimated by multiplying β as read from the curve by $\beta^{0.36}$ and multiplying the resulting time by $\rho_L \lambda / \lambda_{H_2O}$ where ρ_L is in grams per cubic centimeter. If the drop temperature differs appreciably from 130° F., the time may be corrected by a factor $(t_a - 130)/(t_a - t_{drop})$. These curves are considered to be applicable for drop sizes up to about 600μ.

The curves of Figure 5 are applicable to the evaporation of pure liquid drops. They apply to the evaporation from drops containing solids in solution or suspension only during the period when the drop surface is completely wet, as discussed by Ranz and Marshall (10). The period of evaporation when the drops become rigid and heat up is an important fraction of the total drying time. Methods for estimating this fraction of the time require further development.

VARIABLES AFFECTING SPRAY-DRIED PRODUCT PROPERTIES

The dry product from spray drying is particulate and may consist of fine particles ($< 40\mu$) or larger particles (up to 600μ). The particles are generally

spherical and in most cases hollow or porous. It is evident, therefore, that spray drying produces a particulate material with certain desirable characteristics not readily obtainable by other dry-

ing methods. However, properties of the product particles are subject to variations caused by (1) differences in the physical and chemical properties of the feed material itself, (2) variations in feed concentration, (3) variations in feed temperature, (4) variations in air temperature, and (5) variations due to methods and conditions of atomization. Some of the product properties of principal concern in spray-drying applications are (1) bulk density and particle density, (2) bulk appearance, (3) particle size and distribution (this includes dustiness), (3) particle friability or resistance to fracture, (4) moisture content, (5) dispersibility or tendency to agglomerate, and (6) retention of desir-

able product qualities, such as flavor, food value, aroma, color strength, potency, reconstitutability, etc. Although all these properties may not assume equal importance in every case, frequently as many as three of them may have the same degree of importance in a given problem.

Very little theoretical consideration has been given to the manner wherein these product properties can be varied or how they are influenced by the operating variables indicated above. The greatest attention has been given to the bulk density, because of the importance of this variable in determining packaging and handling costs. The property of particle hollowness has also received some attention, but the explanations for this phenomenon have been incomplete and in some cases obscure.

The particle size obtained in spray drying is affected by at least four factors, namely, (1) the method of atomization, (2) the properties of the product itself which may contribute to hollowness, (3) the solids content of the feed, and (4) the drying conditions. The methods of atomization require separate study and are currently being considered by many investigators.

The effect of the properties of the material itself—e.g., whether it is amorphous, colloidal, granular, etc.—on the particle size and shape has received only qualitative consideration.

The effect of solids content of the feed on the dry-particle size can be theorized as follows:

Since the dry-solids content of a drop will remain constant, a weight balance between the initial drop size, as produced by the atomizer, and the final
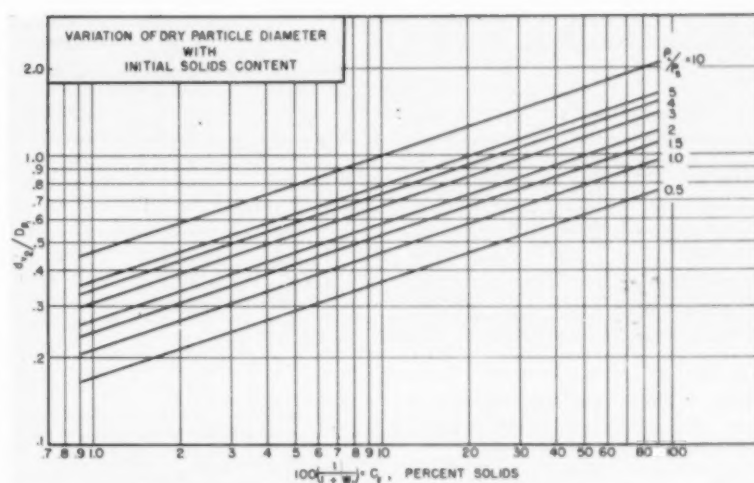
Fig. 6. Effect of solids content and dry bulk density on dry particle size.

drop size of the dry particle may be written as

$$\frac{1}{6}\pi D_{p1}{}^{3}\rho_L\left(\frac{1}{1+W_1}\right)$$
$$=\frac{1}{6}\pi d_{v2}{}^{3}\rho_2\left(\frac{1}{1+W_2}\right) \quad (11)$$

Solving for the ratio of final to initial drop diameter,

$$\frac{d_{v2}}{D_{p1}}=\sqrt[3]{\frac{\rho_L(1)(1+W_2)}{\rho_2(1+W_1)(1)}}, \quad (12)$$

it becomes apparent that the size of the dry particle may be smaller or larger than the initial liquid drop from the atomizer, depending on the final particle

density, ρ_2, and the initial solids content, $(1)/(1+W_1)$. In general, $(1+W_2)/(1)$ will be nearly unity at low final moisture contents. If the value of ρ_2 as a function of drying conditions could be predicted, then the prediction of the variation in initial droplet diameter due to evaporation and drying might also be predicted. However, no procedure for such a prediction has as yet been proposed.

Figure 6, based on Equation (12), shows a plot of the ratio of the final to the initial drop diameter vs. the initial solids content for ratios of ρ_L/ρ_2 ranging from 0.5 to 10. It is of interest to note that for very low initial solids contents, on the order of 1%, the initial drop diameter will be decreased 75% by evaporation only.

Results of Studies of Jet Breakup

As an aid to the interpretation of the drop sizes obtained in the drying studies, drop-size distribution was studied as a function of nozzle diameter, feed rate, and viscosity. Drops were photographed below the point of breakup, and drop images were measured on projections of the negative. This method had the advantage that no physical object was placed in the drop stream, but also had the disadvantage that drops smaller than about 120μ were difficult to distinguish. The photographs were taken with the nozzle mounted on the dryer framework, external to the dryer, where it was subject to the same disturbances as were encountered in the drying tests.

The jet viscosity was varied over a forty fold range by the use of sucrose solutions, the densities varying between 0.996 and 1.269. The nozzle diameters used were 300, 215, 175, and 145μ, and the feed rate varied from 3.12 to 18.87

TABLE 1.—DROP-SIZE-DISTRIBUTION DATA FOR BREAKUP OF VISCOUS JET

Run	D_N, μ	Viscosity, centipoise	F, cc./min.	N_{Re}	D_G, μ	Std. dev.
c	215	4.56	9.42	234	392	1.259
e	300	4.56	18.87	342	464	1.145
f	300	15.4	18.87	106	479	1.123
g	300	15.4	9.42	53	585	1.248
h	215	15.4	6.25	49	478	1.282
i	215	15.4	3.12	25	627	1.320
j	175	15.4	3.12	30	540	1.304
k	175	15.4	6.25	61	367	1.375
l	145	16.2	3.12	35	408	1.250
m	300	36.0	9.40	23	647	1.190
n	300	0.836	18.87	1590	432	1.190
o	215	0.836	18.87	2220	303	1.248
p	215	0.836	9.92	1110	373	1.196
q	215	0.836	6.25	550	392	1.218
r	215	0.836	3.12	225	500	1.350
s	175	0.836	6.25	904	370	1.324
t	175	0.836	3.12	452	360	1.278
u	175	0.836	11.00	1591	302	1.192
v	145	0.836	6.25	1088	276	1.207
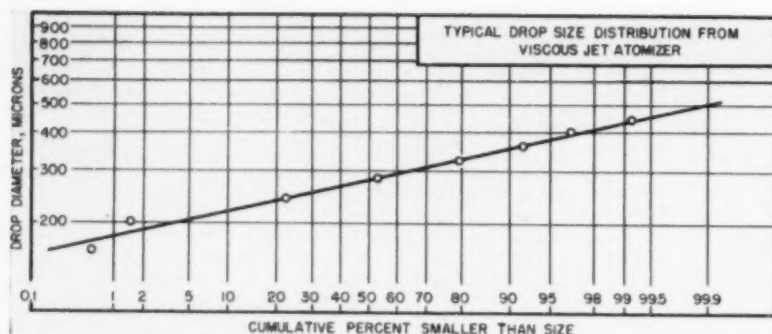
Fig. 7. Log-normal probability plot of drop-size distribution obtained from jet breakup by the Rayleigh mechanism.

cc./min. These variables combined to give Reynolds numbers in the capillaries ranging from 23 to 2230.

Geometric mean drop sizes and standard deviations were determined graphically by plotting the distribution data on log. probability paper. Figure 7 shows a typical distribution. The data for all tests on drop-size distribution were plotted in this manner, and the data are summarized in Table 1. The geometric mean drop size was correlated by the following empirical equation:

$$D_a = 36 D_N^{0.56} N_{Re}^{-0.10} \qquad (13)$$

The experimental data were plotted satisfactorily according to this correlation, as shown in Figure 8. The maximum deviation from the correlation line was 18%, and the average was 7.3%.

In this work vibrations of varying and discontinuous ranges of wave lengths, varying in amplitudes and acting in various directions, were undoubtedly impressed on the jets. Therefore, although Rayleigh's theory held approximately, a range of drop sizes resulted, as Figure 7 indicates. For a given set of operating conditions it was noted that there was a fundamental drop size and that well-defined groups of drops occurred in which the drop diameters were two, three, or more times that of the fundamental size. High-speed photographs of jet breakup showed that impressed vibrations caused marked changes in the uniformity of the drops produced compared with the case when no vibrations other than uncontrolled ones were involved.

HIGH-SPEED MOTION PICTURES OF JET BREAKUP

Motion pictures of jet breakup were taken at a film speed of 7,000 frames/sec. Several interesting phenomena were observed in these pictures, as indicated in the selected frames of Figure 9. After formation, the drop oscillated from an elongated cylinder shape to a flat disk shape. The drops seldom recombined or coalesced on collision, and they usually bounced away from one
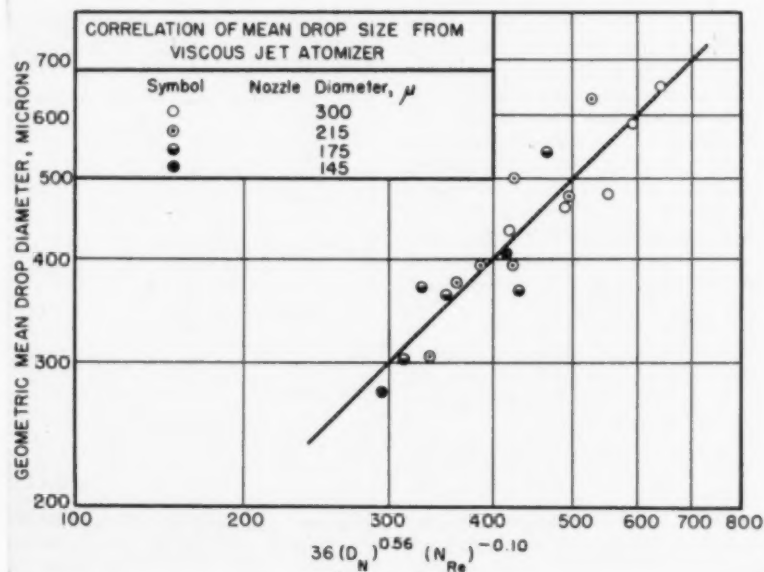
Fig. 8. Experimental values of mean drop size compared with values calculated from Equation (13).

another after collision.

The oscillations which the drops underwent before they attained a stable spherical shape explained how, in some spray-drying and cooling operations, disks or dish-shaped particles can be formed if drying or solidification occurs before sphericity is attained. Further, the mechanism of jet breakup also shows why elongated or dumbbell-shaped particles are sometimes obtained in drying operations when surface drying or case-hardening "sets" ligaments or jets as they are about to break up but cannot complete the collapsing process.

Approximate measurements from these pictures indicated an average ratio of disturbance wave length to jet diameter of about 4.6, based on drop spacing and ignoring the satellite drops. This compares well with Rayleigh's theoretical value of 4.51.

The results of drying studies will be reported in Part II.

Notation

C_f = solids concentration of feed, weight %

D_a = geometric mean drop diameter, μ

D_j = jet diameter

D_N = nozzle diameter

D_p = drop diameter

d_v = volumetric mean dried-particle diameter

F = feed rate, cc./min.

g = gravitational constant

h = film-heat-transfer coefficient on surface of drop

k_f = thermal conductivity of film

N_{Pr} = Prandtl number

N_{Re} = Reynolds number

t_a = air temperature, ° C.

T_a = air temperature, ° K.

t_f = feed temperature, ° C.

t_1 = inlet-drying-air temperature, ° C.

t_2 = exit-drying-air temperature, ° C.

Δt = driving force for heat transfer

u_t = terminal velocity of drop

w = moisture content, weight %, wet basis

W = moisture content, dry basis, weight fraction

β = a parameter [Equation (7)]

λ = wave length of the disturbance on a jet; also latent heat of evaporation

μ = liquid viscosity; also micron

ρ_L = density of liquid drops

ρ_Z = density of dry particle

ρ_a = density of air

ρ_{B1} = bulk density as measured before settling, g./cc.

ρ_{B2} = bulk density as measured after settling, g./cc.

ρ_{BD} = bulk density corrected to zero moisture content, g./cc.

θ = time

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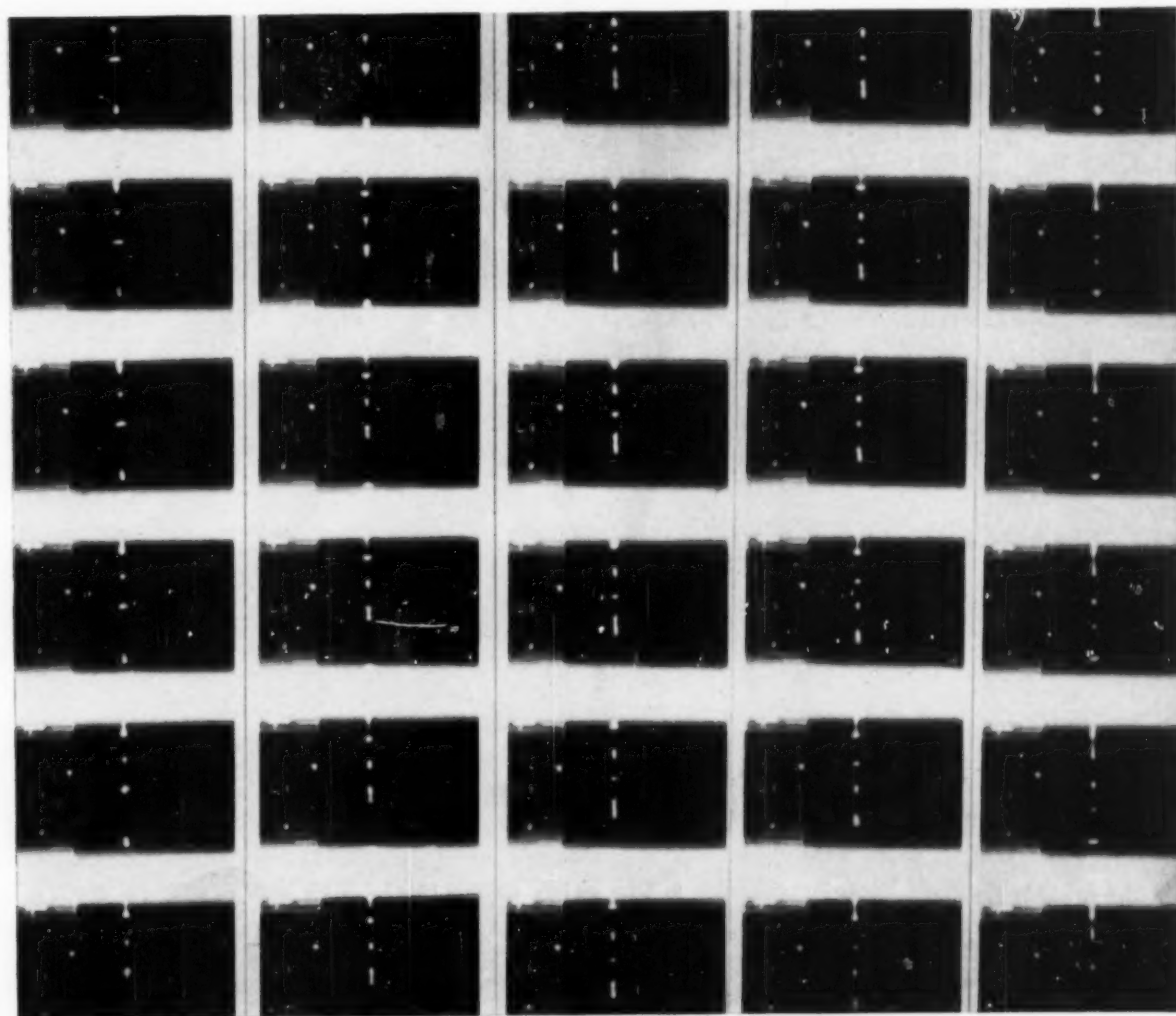
Fig. 9. Successive frames from high-speed motion picture of jet breakup. Scale div. = 1 mm. Nozzle diam. = 890μ., feed rate = 21 cc./min., film speed = 7,000 frames/sec.

the engineer's stake and status in

HUMAN RELATIONS

R. L. Demmerle

Dealing with people is not an exact science, like physics, nor a ritualistic art, like law. It is a constant human endeavor knowing neither total success nor total failure. It is living itself.

This responsibility to mesh one's own personality with those of family, friends, and colleagues is common to all. And a man should consider himself lucky indeed that he is generally more liked than disliked, more accepted than rejected when he realizes the tremendous damage to human relationships that can be wrought by an ill-considered word, action, or attitude.

It is a daily challenge to inspire the confidence and secure the cooperation of others. But it becomes more demanding on those persons charged with the responsibility of directing people while maintaining their respect and loyalty. It is no wonder then that as our industrial civilization matured, proficiency in dealing with people became recognized as a far more important criterion of executive ability than capabilities in economics or even in the basic technology that made industry possible.

The chemical industry proved no exception. Within the lifetime of today's youngest chemical engineers the industry has become a major component of the economy. It is now a big employer, responsible to thousands of stockholders and vitally concerned with the need to maintain and extend its good will with a consuming and investing public. As a result, its executives have had to develop their understanding in the field of human relations at a rate comparable to the expansion of the industry itself.

R. L. Demmerle, assistant director of administration of the General Aniline & Film Corp., is the author of this article which introduced the Symposium on Human Relations at the A.I.Ch.E. meeting in Cleveland. Mr. Demmerle was chairman of that symposium.

Unfortunately many technical graduates of the past two decades were denied the opportunity to develop the broad, easy knowledge of people prerequisite to success in human relations. The depression years and the war that followed made it necessary for them to stress specialized training as the best assurance of security.

This penchant for specialization was encouraged and regarded by growing industry as small close-knit companies now became large departmentalized organizations. Off-the-job as well as on-the-job, specialists tended to build their lives about their vocational groups. Perhaps this was the best way to preserve some degree of personal identity in the huge industrial ant hill. Considerable credit, however, should be given those technically trained men who bucked this trend and maintained an interest in the activities and thinking of people of different trainings and backgrounds.

Objectivity Can Hinder

Another obstacle that many a technical man faces in developing an understanding of people is certain habits of thinking and attitudes developed during his training which inhibit his approach to a human situation. One of the most troublesome of these habits is the blind worship of objectivity. To make him a more competent technologist, he was taught that he must ever guard against any inclination to interject his own attitudes into the interpretation of his data. No doubt this is an effective principle to guide the observation of inanimate matter. But it does not work in the field of human relations. For the observer, no matter how removed from the situation he may think himself, must sooner or later interpret his observations in the light of his own background and feelings.

The statistical findings of the psychologist about a man's personality traits or job aptitudes, for instance, have meaning and utility only when all parties agree on the working definitions of these terms. Even the lie-detector technique, which attempts to correlate measurable physiological data with the truth of a subject's statements, ultimately must depend upon a subjective definition of just what truth is in a particular situation.

Subjectivity has not prevented psychology and other social sciences from being useful to humanity. On the contrary, one of the greatest values of these sciences is to show the individual through his own eyes just how he rates and is rated by the rest of society.

Ego All-Important

This, of course, brings up the importance of personal and group egos, an-

other frequent blindspot in the technical man's understanding of people and their motives, hopes, and fears. There is no point in entering into any human situation, from matrimony to management, without respect for the other person's ego. There is no individual who does not feel that his existence is important and his problems unique. Real humility —not the professional kind, which is just another form of egocentricity—is the willingness not to eliminate but to subordinate the individual ego in the service of a broader cause. All the truly great leaders of history have been able to make others aspire to this type of humility.

To work with other egos, however, requires a clever estimate of their force and direction. This is especially important in any attempt to win the support of others through argument. It is here that the engineer makes the mistake, common to most technical specialists, of assuming that nontechnical people share his interest in how an operation works or how a situation came into being. Actually they may be more interested in the fact that an operation does work or that a situation does exist and most interested in any significance for their own futures.

The engineer, however, because of his worship of documentation, often starts an explanation back at a first cause in the dim past. He carries his listeners forward in a plodding manner, punctuated with side excursions to review approaches that were tried and failed. Eventually he arrives at the point where he can present his conclusion to a weary audience. And frequently he presents alternate conclusions which still require a decision before action can be taken. In brief, he commits the error of not sizing up the impatient self-interest of his audience and tailoring his presentation to fit. Moreover, if his audience consists of his top management group, he brands himself as an intellectually torn person, incapable of determining a course of positive action.

This approach is very different from that of the promotional-minded executive, who invariably offers his suggested course of action early in a presentation, while the audience is fresh, interested, and receptive. The remaining 99 per cent of his discussion is then devoted to arguments in favor of his plan. If he loses any of his audience along the way, it doesn't matter too much; the punch line has been stated.

People—Constant Variables

The engineer's preference for step-by-step thinking and conversation probably springs from his passion for what he terms a systematic approach to living.

Every technical man nurses a desire to make the world more orderly. It is no wonder that technocracy, preaching one of the most systematic approaches to basic economics ever proposed, had a high percentage of engineers among its founders and followers. The technocrats' proposal to place currency on a backing of energy and work units rather than the arbitrary standard of gold was ingenious. But more than that, it was characteristic of the engineer's distrust of nonabsolute standards. To the technical man, almost anything which cannot be placed upon a measurable, predictable basis is subject to doubt and suspicion.

This attitude, born during his training and nurtured by his professional work, is perhaps the biggest single barrier the engineer faces in his attempt to understand people and their loves, fears or desires. When confronted with a problem in human relations, the technically trained man instinctively seeks some factor in it that can be measured by absolute standards. There, he feels, the solution lies. If no such factor is found, he will usually decide that the problem is incapable of logical resolution.

It is amusing to observe how often the engineer will reject the use of relative standards and an empirical approach to human situations yet readily embrace them in developing certain aspects of his own technical work not yet fully explainable. It would almost seem that he can more readily tolerate nonabsolutes in physical data than in human psychology. But until knowledge of the human personality becomes much more clearly defined, the empirical approach offers the best practical means of understanding people.

The biggest advantage of this practice is that it permits the acceptance of some of the so-called "inconsistencies" of human nature. Moreover it makes it possible for the individual to recognize them in himself and work with them in others. In many cases, it even offers an explanation of why some two-plus-two situations add up to three or five.

Serving Two Masters

Any empirical approach to human behavior is helped by an appreciation of the struggle that goes on within all of us, the constant tug of war between a desire for security and an urge to venture. The dilemma is a deep one. The security involved takes in much more than an assured source of livelihood and a roof over one's head and a pension after one reaches sixty-five. It also includes a wish to conform, a desire for tradition, and a hunger for approval. Security in this sense means all those things that make tomorrow more predictable on the basis of today.

On the other side of the balance is venture, which includes taking a chance, but much more too. It encompasses all those drives that push us toward being different, asserting ourselves, and in general trying to make sure that the world will always remember that we were once part of it.

It would be easy if everyone could be classified into one or the other of these extreme types. But, fortunately for the sake of happiness, the great bulk of humanity falls somewhere in between. Some severe personality disorders are characterized by either an obsessive desire for security or a passion for accelerated change. The normal person, however, makes an attempt to work out a life that combines the best portions, in his eyes, of the security that is afforded by constancy and the excitement that is offered by change. In his daily living he will usually drift toward those persons who maintain a similar ratio.

Paradoxy enters the picture, however, when the hidden heroes of the quiet people are examined. The $4,000 a year bookkeeper who thrills to the emotional action of baseball will argue every spring that his $75,000 a year big-league idol is underpaid. And the plain little housewife, married to the same man for twenty years, will worry vicariously about whether her favorite screen actress will at last find true happiness in a fifth marriage.

By the same token, those who live obviously exciting lives and would not think of foregoing them frequently envy the constancy of an undramatic existence. There is always that persistent worry in individuals of this type that perhaps they may be missing something.

Although these may be extreme examples, they illustrate what the man skilled in the art of human relations has always known: that people live half the time in the world as they see it and the other half in the world as they wish it were. As a consequence, the individual in his loneliness will seek, trust, and often emulate the person who seems to have found a happy reconciliation between these two worlds. For contentment begets contentment, and confidence inspires trust. Every one who would influence others therefore must have a positive attitude toward life. It must be based upon a growing philosophy that he constantly extends by his questions to himself and others.

It is natural to classify people by the ideas they express and the attitudes they demonstrate. But it is necessary to go a little deeper than this, for it is important to find out how people came by their ideas and why their attitudes developed in the form they did. This, of course, takes time and patience and most of all a willingness to appreciate the role that

the other peron's emotional pattern—his likes and dislikes—plays in forming that person's attitudes.

This seems to be the point where many academically minded people fail in human relations. They try to talk around —or talk other people out of—emotionally based attitudes. They like to believe that once the emotional factors are removed from the scene, all human problems will respond to easy solution. But if emotion, like water, is part of the world in which we live, it would seem a better course to learn to swim than attempt to dry up the oceans.

The man versed in the practical side of human relations accepts emotional bias and tries to work with it, knowing that no one is sufficiently mature to separate the emotional from the reasonable components of every attitude he develops. A man's fancies, fetishes, and fears are very real to him and provide paths by which he may be approached. Admittedly, the opportunist may use them to exploit him. But the bigger man will recognize them as an opportunity to create an atmosphere of mutual understanding.

The Engineer's Advantages

If the engineer can throw off his prejudice against emotional elements, he will find that on many scores he is in an enviable position to inspire trust and confidence. For one thing, he is by nature a constructive person, not given to exploiting others. His profession represents the very antithesis of parasitical living, thus earning him the respect of hardworking people. A second point in his favor is that people generally have a deep respect for men who attempt to extend the boundaries of human knowledge. Each advance seems to bring greater security and comfort with it and often makes life more exciting and interesting, thus serving the previously mentioned human drives for security and change.

The engineer's reputation for honesty is another factor that works in his behalf in human relations. Unlike the politician, or for that matter the salesman, the word of the engineer is rarely questioned. In fact, he ranks with the physician as an accredited witness in court. Finally, the engineer enjoys a prestige that accrues to people who work with abstract concepts. And as long as he uses his ability in this direction to simplify, not complicate, his relationships with other persons, this prestige will work to his favor in human relations.

All these advantages are real and strong. But they avail the technical man little unless he accepts them for what they are—a few valuable properties in the big play of life, getting along with people.

Resistance and Compressibility of Filter Cakes

Part III: Under Conditions of Centrifugal Filtration

H. P. Grace

E. I. du Pont de Nemours and Company, Wilmington, Delaware

Interpretation of compression-permeability data in terms of conditions of centrifugal filtration is considered for general and simplified cases. Forces existing in centrifugal filtration are analyzed, and a simplified centrifugal-rate equation is developed. Results of centrifugal-drainage rate and centrifugal-filtration runs for four materials are correlated in terms of this simplified-rate equation. Values of average specific cake resistance for centrifugal filtration are compared for two materials with values predicted from compression-permeability data and values determined by pressure filtration. Inherent difficulties in the sizing of centrifugal filters for semicompressible materials are considered, and shortcomings of methods now used are analyzed.

Previous parts of this paper have covered the study of filter-cake properties under conditions of both uniform compressive pressure and actual pressure filtration. This further part presents a logical extension of the same approach to the more complicated case of centrifugal filtration.

A number of factors exist in both theory and practice which make the sizing of a perforate-basket centrifuge more difficult and less understood than the sizing of conventional vacuum or pressure filters. The geometry of the centrifuge bowl results in an effective filtration area which decreases as cake thickness increases, and this results in a change in superficial fluid velocity through cake. The driving force causing filtration not only is the Δp across the cake but also includes the centrifugal force on the liquid as it flows through the cake. The centrifugal field is not uniform as in gravity filtration but varies considerably with bowl radius. The compressive pressure acting on cake solids at any point in the cake results not only from friction loss of the fluid flowing, but also from action of the centrifugal field on the mass of cake solids and from kinetic-energy change in the fluid flowing through the cake. On the practical side, conditions in a plant-size centrifuge cannot be duplicated in the usual laboratory-size perforate-basket centrifuge; extrapolation to both higher compressive pressures and thicker cakes is necessary. Also, in operation of plant-size machines, a $\frac{1}{2}$- to $3\frac{1}{2}$-in. heel of cake must be left at the end of each cycle if a mechanical unloading plow is used, and this results in a relatively high initial filtration resistance at the beginning of each subse-

Part III is the final section of this paper. Parts I and II appeared in the June and July issues.

quent cycle, since this heel has been subjected to the maximum compressive pressure developed in the centrifuge.

Considerable difficulty has been encountered in the sizing of perforate-basket centrifuges for filtering very fine crystalline materials. Sizing has been done most often from small-scale tests in a conventional 12-in.-basket centrifuge run at a speed to give the same centrifugal force at the basket wall as was anticipated in the plant machine. This involves extrapolation of results with 1.5-in.-thick cakes to the 3- to 5-in. cakes expected in plant machines. The specification of undersize machines has resulted in certain cases, probably because the extrapolation did not take into account the highly compressible nature of the cakes involved and the increased compressive pressures developed with thicker cakes.

Theory and Rate-Equation Development

Burak and Storrow (1), Inglesent and Storrow (2), Maloney (3), and Smith (4) have proposed modifications of the rate equation for pressure filtration in order to make it applicable to centrifugal filtration. These proposed modifications do not agree in their definition of the driving force in centrifugal filtration corresponding to the Δp term

in pressure filtration. Also, the proposed modifications are applicable only to the simplified case of a nearly incompressible cake, and they neglect the filter-medium resistance entirely. The approach suggested by Burak and Storrow appears reasonable, but these authors were not able to verify their proposed relationship experimentally (1). The proposed relationships of Maloney and of Smith have also never been verified. Although unpublished at this writing, further work by Storrow and associates is believed to have experimentally verified their proposed relationship for several materials. This relationship, as derived by Burak and Storrow, was in the form

$$Q = \frac{4\pi^3 N^2 h K_e}{\mu g} \cdot \frac{(r_o^2 - r_f^2)}{\ln (r_o/r_e)} \qquad (1)$$

where K_e is the average permeability of the cake.

An analysis of factors involved in centrifugal filtration is presented in the Appendix. The differential pressure balance at any instant in the filtration takes the form:

$$-dp = -\frac{\rho_f \omega^2}{g_e} r dr + \frac{\rho_f}{g_e} \left(\frac{Q}{2\pi h}\right)^2 \frac{dr}{r^3} + \frac{Q \rho_s \mu}{2\pi h g_e} a_p (1 - \epsilon_p) \frac{dr}{r} . \qquad (2)$$

Fig. 1. Shallow-bowl centrifugal filter.

sistance a for centrifugal filtration directly from compression-permeability data, as in simple pressure filtration, thus eliminating the need for any actual centrifugal filtration measurements in solution of a problem. However, the final relationship developed cannot be numerically integrated without a knowledge of a_p or p_s as a function of bowl radius r. This expression, as derived in the Appendix, is

$$\int_0^{p_s} \frac{dp_s}{1-\epsilon_p} = \frac{(\rho_s - \rho_f)\omega^2}{g_c} \int_{r_c}^{r_o} r \, dr$$

$$+ \frac{Q\rho_s}{2\pi h g_c} \int_{r_c}^{r_o} a_p \frac{dr}{r}. \qquad (4)$$

Thus, if a_p is assumed to be some particular function of r, this can be integrated numerically with compression-permeability data to obtain the filtration rate Q. However, it appears that $a_p = f(r)$ may depend on the form of the $a_p = f(p_s)$ relationship and therefore be different for each material, thus preventing a general solution. For the special case when r_o is relatively large compared to the cake thickness $r_o - r_c$ and when dp_w is small compared with dp_f, Equation (4) simplifies to the form

$$\int_0^{\Delta p'_c} \frac{dp_s}{(1-\epsilon_p)a_p} = \frac{\Delta p'_c}{(1-\epsilon)a}, \quad (5)$$

as derived in the Appendix. In this case the over-all driving force $\Delta p'_c$, causing liquid flow through the cake, can be expressed as

$$\Delta p'_c = (p_2 - p_1) + \frac{\rho_f \omega^2 (r_o^2 - r_c^2)}{2g_c}. \qquad (6)$$

For this simplified case, Equation (5) can be solved directly for $(1-\epsilon)a$ using numerical integration of compression-permeability data.

Experimental Equipment and Procedure

To study the effect on filtration rate and cake properties of filtering under an actual centrifugal field, the special shallow-bowl centrifuge shown in Figure 1 was designed and built. The bowl was 13-in. I.D. by 2-in. inside depth, with a ¾-in.-thick Lucite acrylic resin top ring fabricated to permit the build-up of a 4½-in.-thick cake. The top ring was etched with 1-in. ring markers and at 90° radii was etched with $\frac{1}{16}$-in. markers between rings. This permitted constant observation and control of cake thickness and liquid level in the centrifuge during operation. The woven filter medium was supported on 16-over 8-mesh screen over the perforated-basket wall. The centrifuge was driven by a 3-hp. variable-speed Thymotrol unit with 0.5% speed regulation. A speed range of 0 to 3400 rev./min. (0 to

This states that the pressure drop across an increment of cake equals the sum of the centrifugally induced hydraulic head across the increment, the change in kinetic energy of the liquid flow through the increment, and the friction loss suffered by liquid flowing across the increment. In general, a solution is not possible unless the variation of a_p and ϵ_p with radius r is known. However, for nearly incompressible materials where variation in a_p and ϵ_p is small, a simplified solution can be obtained, as shown in the Appendix, in terms of a and ϵ. The over-all rate equation can then be written as

$$Q = \frac{4\pi^3 N^2 \rho_f h (r_o^2 - r_r^2)}{\mu \left[\rho_s a (1-\epsilon) \ln \frac{r_o}{r_c} + \frac{R_m}{r_o} \right]}. \qquad (3)$$

This simplified rate equation for centrifugal filtration has the same general form as the pressure-filtration-rate equation. Thus, a plot of $1/Q$ vs. $\ln (r_o/r_c)$, for a centrifugal filtration run with constant speed N and liquid level r_f should yield a straight line with slope proportional to a and an intercept proportional to R_m/r_o provided the cake behavior can be approximated by the assumption of an incompressible cake. This simplified rate equation is similar to that proposed by Burak and Storrow (1) except that the filter-medium resistance is included, and a, average cake resistance, is included rather than average cake permeability K_c.

The total compressive pressure acting in centrifugal filtration results from the friction loss through cake and from the net weight of dry solids supported in the suspending liquid and acted upon by the centrifugal field. It had been hoped that analysis would permit calculation of the average specific cake re-

2,000 g) was provided. A tachometer generator provided speed indication, but final speed measurements were made with a GR Strobotac tachometer. Feed suspension or filtrate fed to the centrifuge bowl passed through a small centrifugal pump with Variac speed control and was distributed within the bowl by a Bunsen burner wing tip on the feed-pipe outlet. This was oriented to give uniform feed distribution over the depth of the bowl and uniform cake thickness over the 2-in. bowl depth.

Experimental Procedure

After a number of preliminary runs with Standard Super-Cel and G-60 Darco, which were made to check operating procedures, a blank run was made with only the used N.F.M. No. 10 cotton twill cloth in place. Filtrate was passed through the cloth at various centrifuge speeds and with various liquid heads maintained over the cloth by controlling feed rate with Variac and pump. Filtrate-discharge rates were determined by weighing filtrate collected for timed periods. Filtrate was continuously recirculated through the centrifuge by pump. Results were correlated satisfactorily by a logarithmic plot of cloth resistance R_m vs. pressure drop across the cloth, $(p_1 - p_o)$, where

$$(p_1 - p_o) = \frac{\rho_f \omega^2}{2 g_c} (r_o^2 - r_f^2).$$

The slope of this plot (Fig. 2) showed the filter cloth to be slightly compressible in the pressure-drop region covered.

Later two types of experimental runs were made with several filter-cake materials. The first was termed a "drainage-rate" run. In this type, suspension containing a known weight of dry solids was introduced into the bowl and a cake deposited at a constant low speed while a constant r_f was maintained by recirculating filtrate. When the cake was deposited, rate of filtrate flow and r_c were noted. Centrifuge speed N was then increased stepwise, holding r_f constant and determining filtrate rate and r_c at each speed. A range of 400 to 2500 rev./min. was usually covered. However, difficulties were encountered when runs of this type were attempted with R-110 titanium dioxide or BW-200 Solka Floc; in these cases radial cracks developed in the cakes as a result of compression as speed increased, and valid results could not be obtained with these materials. Runs yielding consistent data were obtained with Standard Super-Cel and G-60 Darco. Typical data and calculations for such a run are shown in Table 1.

The second type of experimental run was termed a "centrifugal-filtration run" although it was actually a combination centrifugal-filtration-drainage-rate run. This type of run was completed at a constant speed and repeated for various centrifuge speeds of from 400 to 2000 rev./min. In making the run, r_f was held constant by filtrate recirculation and the cake was deposited stepwise by addition of various slurry volumes containing known weights of dry solids. After deposition of each cake increment, filtrate drainage rate and cake thickness were noted and the next increment of cake was formed. Cake
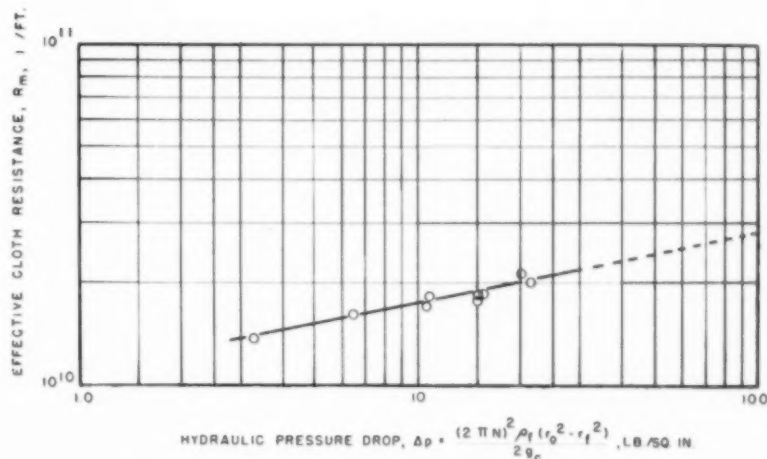
Fig. 2. Resistance of filter medium in centrifugal filtration.

thicknesses up to $3\frac{1}{2}$ in. were formed by this method in making these runs. Since r_f and speed were maintained constant in these runs, the driving force causing filtration remained essentially constant (neglecting kinetic energy changes), and these runs approximate constant-pressure-drop filtration runs. Typical data and calculations for such a run are shown in Table 2.

Experimental Results

The results of drainage rate runs on Standard Super-Cel, G-60 Darco, and R-110 titanium dioxide suspensions are shown in Figure 3. The filter cloth alone and Standard Super-Cel show straight-line relationship, but with a slope of 1.7 where a slope of 2 would have been expected for an incompressible material. The results for the G-60 Darco show an initially constant slope of about 1.7. At higher speed, the slope decreased to about 0.7 at 2,000 rev./min., indicating increasing cake compression. With the titanium dioxide an initial slope of 1.5 was obtained, but the cake then developed radial cracks which caused a sharp rise in slope, while the slope dropped rapidly at high speed as the cracks closed under further compression. Difficulty with cake cracking as speed was increased with both titanium dioxide and BW-200 Solka Floc placed the results obtained by this method on a questionable basis. The method was abandoned in favor of the constant-speed method of making centrifugal-filtration runs.

The results of a series of such centrifugal-filtration runs with Standard Super-Cel suspensions are shown correlated by two methods in Figure 4. The left-hand plot shows correlation by means of a rearranged form of the simplified rate equation [Equation (5)]. Straight-line correlations result, and α and R_m can then be calculated from the slope and the y-axis intercept. Several

runs appear to have a negative y intercept, however, which is difficult to explain since no initial bleed occurred but which may represent only experimental error, as the value of y intercept is small in any case. The right-hand plot of Figure 4 shows curves of experimental data plotted in the usual manner of pressure filtration; however, centrifuge bowl area A_c has been used in one case, while logarithmic mean cake area A_{lm} has been used in both the ordinate and the abscissa terms in a second case. Use of A_c resulted in curved plots, but use of A_{lm} resulted in essentially straight-line correlations of the data for runs at various constant speeds. Again calculation of α and R_m from slope and intercept of the straight-line curves was possible; the values of α obtained were in good agreement with those obtained from the previous correlation and use of the simplified rate equation, provided the Δp term of the pressure-filtration equation was replaced by a driving force term $\Delta p'$, which for centrifugal filtration is not only the pressure-drop term but includes the centrifugally induced head throughout the cake.

The results of centrifugal-filtration runs with titanium dioxide, G-60 Darco, and BW-200 Solka Floc are presented in Figure 5 in the same manner as were the results for Standard Super-Cel in Figure 4. Despite the fact that these materials form compressible cakes, as indicated by compression-permeability data, the correlation of results for these three materials is good. This unexpected result is not easily explained, but may result from a combination of factors which tended to minimize the relative increase in compressive pressure which occurred in the course of each of these centrifugal-filtration runs. Thus, in these centrifugal-filtration runs at constant speed with constant liquid head,

TABLE 1.—TYPICAL CENTRIFUGAL-DRAINAGE-RATE DATA

Single 0.89-lb. G-60 Darco cake deposited from 56-g./liter distilled-water suspension at 422 rev./min. centrifuge speed

$A_c = 0.568$ sq.ft. $r_o = 0.541$ ft. $r_f = 0.208$ ft. $h = 0.166$ ft.

$T_{filtrate} = 25.2°$ C., NFM No. 10 cotton-twill cloth (used)

Centrifuge speed, N, rev./min.	N² (rev./sec.)²	Wt. filtrate collected, lb.	Collection time, sec.	Drainage rate, Q_m, lb./sec.	Cake thickness, in.	r_c, ft.	$\ln \frac{r_o}{r_c}$	$1-\epsilon$	Total Δp, lb./sq.in.	R_m*, 1/ft.	R', 1/ft.²	α,** ft./lb.
422	49.5	2.22	178.4	0.0125	1¹⁄₁₆	0.453	0.1765	0.155	3.28	1.48×10^{10}	13.3×10^{10}	3.10×10^{10}
600	100	2.18	100.8	0.0216	1¹⁄₁₆	0.453	0.1765	0.155	6.64	1.60	15.55	3.70
810	182	2.98	85.1	0.0350	1.00	0.458	0.166	0.167	12.0	1.80	17.5	4.10
1010	284	2.69	58.0	0.0464	1.00	0.458	0.166	0.167	18.8	1.97	20.6	4.91
1300	470	2.20	35.1	0.0627	1.00	0.458	0.166	0.167	31.1	2.20	26.2	6.13
1590	705	2.99	39.1	0.0765	1⁵⁄₁₆	0.464	0.1555	0.175	46.6	2.40	31.0	7.88
1920	1025	2.24	24.0	0.0933	1⁵⁄₁₆	0.464	0.1555	0.175	67.8	2.57	36.9	9.52
2102	1230	2.31	23.6	0.0978	⅞	0.468	0.1442	0.188	81.3	2.67×10^{10}	42.3×10^{10}	11.05×10^{10}

* Taken from R_m vs. Δp curve for blank run, no cake.
** Corrected for cloth resistance as indicated below

$$\text{Total } \Delta p' = \frac{(2\pi N)^2 \rho_f (r_o^2 - r_f^2)}{2g_c} = \frac{(19.7)(62.3)(0.294 - 0.043)N^2}{(32.2)(144)} = 0.0664 \, N^2 \qquad \rho_c = \frac{0.89}{\pi(0.294 - r_c^2)(0.166)} \qquad 1 - \epsilon = \frac{\rho_c}{124.6}$$

$$R' = \frac{4\pi^3 h \rho_f (r_o^2 - r_f^2)N^2}{\mu Q_m} = \frac{33.6(10)^6 N^2}{Q_m} = \frac{1}{\text{ft.}^2}$$

$$R' = \rho_c (1-\epsilon)\alpha \ln \frac{r_o}{r_c} + \frac{R_m}{r_o}$$

$$\alpha = \left(R' - \frac{R_m}{r_o} \right)\left(\frac{1}{\rho_c(1-\epsilon) \ln \frac{r_o}{r_c}} \right) = \left(R' - \frac{R_m}{0.541} \right)\left(\frac{1}{(124.6)(1-\epsilon) \ln \frac{r_o}{r_c}} \right) = \frac{\text{ft.}}{\text{lb.}}$$

TABLE 2.—TYPICAL CENTRIFUGAL-FILTRATION-RUN DATA

Centrifuge speed, $N = 2040$ rev./min. $r_o = 0.541$ ft. $F_o = 765 \times G$ $A_c = 0.568$ sq.ft. $r_f = 0.208$ ft. $T_{filtrate} = 25.5°$ C.

$h = 0.166$ ft. Feed suspension = 55.5 g. Standard Super-Cel/liter distilled water

National Filter Media Co. No. 10 (181-TW) cotton-twill cloth (used)

Total wt. suspension filtered, lb.	Total wt. solids filtered, W, lb.	Wt. filtrate collected, M, lb.	Collection time, θ sec.	Reciprocal rate, $1/Q_m$ sec./lb.	$\Delta\left(\frac{M}{A_c}\right)$ hr./lb./sq.ft.	Cake thickness, in.	r_c, ft.	$\ln \frac{r_o}{r_c}$	W/A_c, lb. solids/sq.ft.	W/A_{lm}, lb. solids/sq.ft.	$\frac{\Delta\theta}{\Delta(M/A_{lm})}$ hr./lb./sq.ft.
13.09	0.69										
16.91	0.89	4.05	47.5	11.55	0.00182	1¼	0.437	0.214	1.57	1.75	0.00162
		4.04	47.2	11.50							
20.55	1.08	4.04	54.4	13.45	0.00212	1½	0.417	0.260	1.90	2.18	0.00185
		4.01	56.5	14.20							
24.01	1.27	4.03	68.6	17.03	0.00269	1¾	0.396	0.313	2.24	2.64	0.00228
27.46	1.45	4.06	78.7	19.35	0.00306	2.0	0.375	0.366	2.55	3.07	0.00264
30.76	1.62	4.05	85.7	21.15	0.00335	2¼	0.354	0.424	2.85	3.51	0.00271
33.96	1.79	3.99	101.0	25.3	0.00400	2½	0.333	0.487	3.15	4.01	0.00314
36.98	1.95	4.01	115.3	28.8	0.00454	2¾	0.312	0.551	3.43	4.50	0.00346
39.85	2.10	4.16	134.2	32.3	0.00510	3.0	0.292	0.618	3.70	4.97	0.00379
42.50	2.24	4.20	149.4	35.5	0.00560	3¼	0.271	0.692	3.90	5.46	0.00400
44.90	2.37	4.18	156.5	37.5	0.00592	3⁷⁄₁₆	0.255	0.748	4.18	5.97	0.00415
47.0	2.48	4.16	169.1	40.7	0.00643	3⅝	0.249	0.787	4.37	6.42	0.00438

Total cake volume $= \pi(0.541^2 - 0.249^2)(0.166) = 0.121$ cu.ft. Avg. $\rho_c = \frac{2.48}{0.121} = 20.5$ lb./cu.ft. $1 - \epsilon = \frac{\rho_c}{\rho_s} = \frac{20.5}{124.6} = 0.165$

$$\text{Total } \Delta p' = \frac{(2\pi N)^2 \rho_f (r_o^2 - r_f^2)}{2g_c} = \frac{(2)(9.85)(1155)(62.3)(0.294 - 0.043)}{(144)(32.2)} = 76.7 \text{ lb./sq.in.}$$

Plot of $1/Q_m$ vs. $\ln \frac{r_o}{r_c}$ gave straight line of slope $= \frac{40 - 1.25}{0.776} = 50.0$ and y intercept $= 1.25$

$$\alpha = \frac{4\pi^3 N^2 h \rho_f (r_o^2 - r_f^2)}{\mu \rho_c (1-\epsilon)} \text{ (slope)} = \frac{(4)(30.9)(1155)(0.166)(62.3)^2(0.294 - 0.043)}{(0.884)(0.000672)(124.6)(0.165)} (50.0) = 9.44 \times 10^{10} \text{ ft./lb.}$$

$$R_m = \frac{4\pi^3 N^2 h \rho_f r_o (r_o^2 - r_f^2)}{\mu} \text{ (intercept)} = \frac{(4)(30.9)(1155)(0.166)(62.3)^2(0.541)(0.294 - 0.043)}{(0.884)(0.000672)} (1.25) = 2.62 \times 10^{10} \text{ 1/ft.}$$

the component of compressive pressure resulting from the friction loss remains practically constant, and the major cause for an increase in compressive pressure and an increase in α (changing slope in correlations of Figures 4 and 5 is an increase in weight of solids deposited as cake. For the case of titanium dioxide, the low flow rates obtained limited the run to relatively thin cakes, and the increase in compressive pressure as cake built up was not more than 20% of the constant compressive pressure resulting from friction loss. Under these conditions only a 5 to 10% increase in α could be expected over the run, and the corresponding change in slope may be masked by the scatter of experimental data. For the case of the G-60 Darco and BW-200 Solka Floc, cakes up to $3\frac{1}{2}$ in. thick were used, but a combination of low ρ_s and high ϵ_p (0.85) resulted in the same small relative increase in compressive pressure. Thus the expected 5 to 10% increase in α and in slope of the correlating curve may again be masked by scatter of the experimental data.

Figure 6 shows comparative values of specific cake resistance for Standard Super-Cel and G-60 Darco as obtained from the various methods of measurement. Values of α_p vs. p_s as obtained from compression-permeability measurements are shown together with a predicted curve of α vs. Δp for pressure filtration and actual values of α vs. Δp for pressure filtration and for centrifugal filtration (for the centrifugal-filtration case Δp was taken as $\Delta p'$, the over-all driving force in centrifugal filtration). For Standard Super-Cel, the values of α obtained by centrifugal filtration fall close to both the predicted and the actual pressure-filtration values over the Δp range covered. For G-60 Darco, the values of α obtained from centrifugal filtration fall close to both the predicted and the actual pressure-filtration values at low values of filtration driving force, but at values of driving force above 50 lb./sq.in. the values for centrifugal filtration rise 20% above those for pressure filtration, indicating additional cake compression of the centrifugal cake at the higher centrifuge speeds.

On a theoretical basis there is strong evidence that with compressive materials a higher value of average cake resistance α results in centrifugal filtration than in pressure filtration under conditions of equal over-all driving force. Only limited direct experimental evidence of this has been obtained. Analysis indicates, however, that the material properties and experimental conditions in the centrifugal-filtration runs with the several compressible cakes may have

Fig. 3. Drainage-rate data for centrifugal filtration

$r_o = 6.5$ in. $r_f = 2.5$ in.

FILTRATE DRAINAGE RATE, Q_m, LB./SEC

CENTRIFUGE SPEED, N, REV./MIN.

● Used filter cloth, no cake
  NFM #10 (181-TW) cotton twill
○ 0.89-lb. cake of Standard Super-Cel
  Deposited from 50-g./l. distilled-water suspension
△ 0.89-lb. cake of G-60 Darco
  Deposited from 50-g./l. 0.01 M $Na_4P_2O_7$ suspension
▲ 1.0-lb. cake of R-110 grade TiO_2
  Deposited from 116-g./l. distilled-water suspension
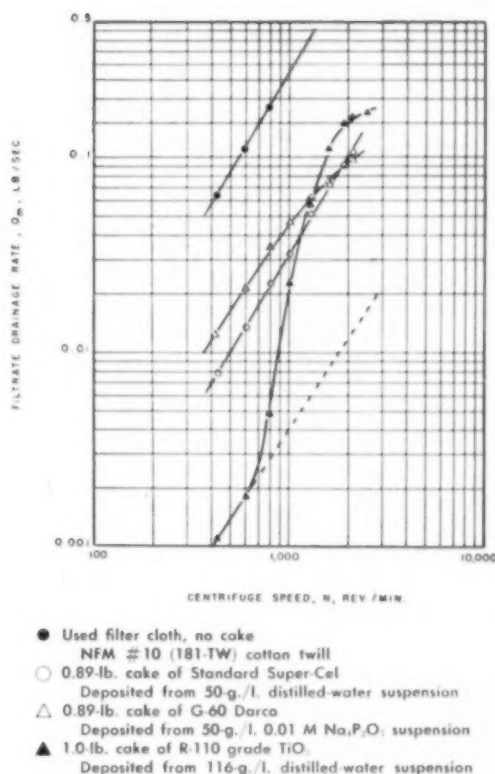
combined to prevent good definition of this effect.

Sizing Procedures for Centrifugal Filtration

In actual plant operation it is usually not practical to operate with more than a slight head of feed suspension or wash liquor over the cake surface. In a deep bowl this liquid head tends to be unstable, even with stabilizing rings in the bowl, and results in unbalance accompanied by bad vibration. For this reason plant operation with a constant liquid level (constant filtration driving force at constant speed) becomes impractical, and sizing on the basis of plant operation with practically no liquid head above the cake becomes the safest practice. For an incompressible cake, this type of operation approximates a constant-rate filtration. If the centrifugal field were uniform and effective filtration area of cake did not change (approximated by centrifuge of very large diameter), a constant-rate filtration would result from this type of operation, and the flow rate would be independent of cake thickness. In an actual plant centrifuge, the filtration rate would decrease with increasing cake thickness, even with an incompressible material, because of both the nonuniform centrifugal field and the decreasing effective flow area of cake. For the case of a

compressible cake, the filtration rate would decrease considerably more because of the increased average specific cake resistance accompanying the increase in compressive pressure, which is developed both from the rising hydraulic driving pressure developed in the cake liquid and from the rising weight of cake solids.

Figure 7 illustrates one of the difficulties inherent in current methods of scaling up centrifugal-filtration data from small to large centrifuges. These methods use a small centrifuge where cake thickness is limited to $1\frac{3}{4}$ in. maximum and are based on operation of both small and large machines at a speed giving the same centrifugal force at the basket wall in both machines. With these limitations, Figure 7 shows that for 1,000-g operation at r_o (the basket wall), the plant-size machine develops an over-all hydraulic driving force $\Delta p'$ of 230 lb./sq.in. with a 4-in. cake thickness, while the small machine develops a driving force of only 50 lb./sq.in. with a $1\frac{3}{4}$-in. cake. For an incompressible cake this poses no problem, and the scale-up can be made on the basis of the average specific cake resistance in the small machine. On the other hand, for a compressible cake, the average specific cake resistance measured in the small machine under the relatively low compressive pressure developed in it must be
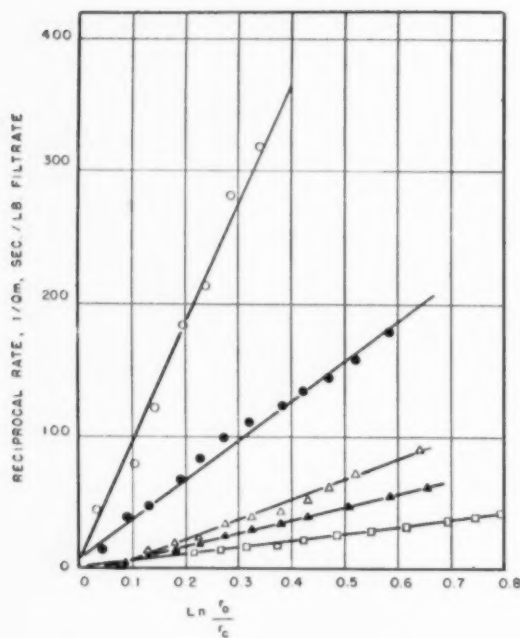
○ N = 417 rev./min.   Δp' = 3.2 lb./sq.in.
● N = 710 rev./min.   Δp' = 9.3 lb./sq.in.
△ N = 1000 rev./min.   Δp' = 18.5 lb./sq.in.
▲ N = 1400 rev./min.   Δp' = 36.2 lb./sq.in.
□ N = 2040 rev./min.   Δp' = 77.2 lb./sq.in.

Fig. 4 (above, left and right). Correlation of constant-pressure centrifugal-filtration data

Standard Super-Cel 55 g./l. of distilled water
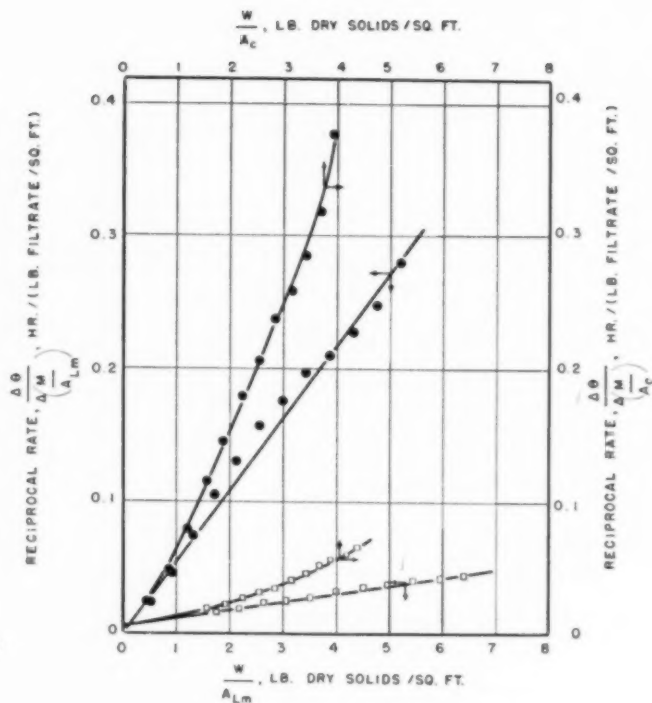
○ R-110 Grade TiO₂, 116 g./l. distilled water
    N = 2030 rev./min.   Δp' = 76.0 lb./sq.in.
● G-60 Darco, 50 g./l., 0.01 M Na₃P₃O₇
    N = 1415 rev./min.   Δp' = 36.7 lb./sq.in.
△ BW-200 Solka-Floc, 42 g./l. distilled water
    N = 1000 rev./min.   Δp' = 18.5 lb./sq.in.

Fig. 5 (below, left and right). Correlation of constant-pressure centrifugal-filtration data
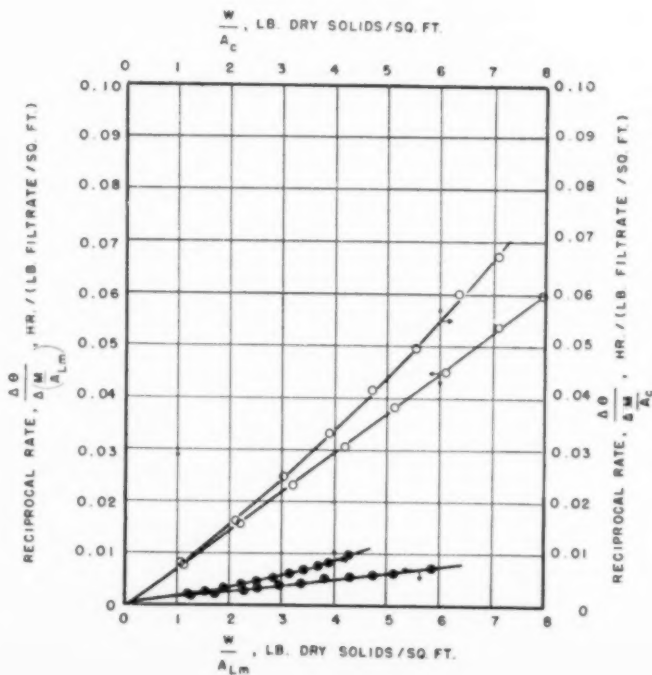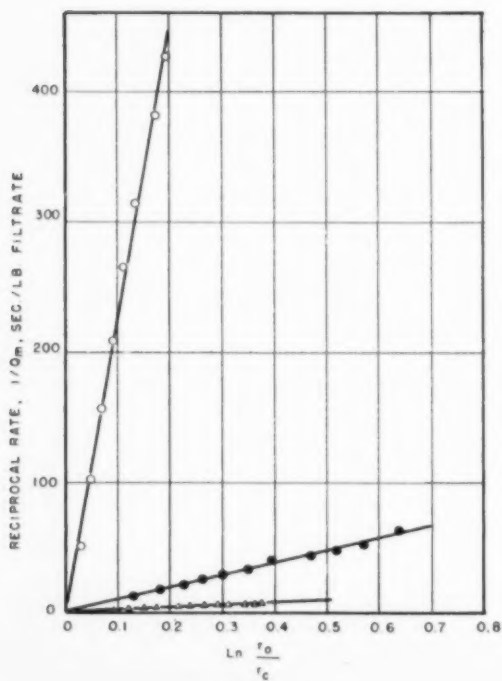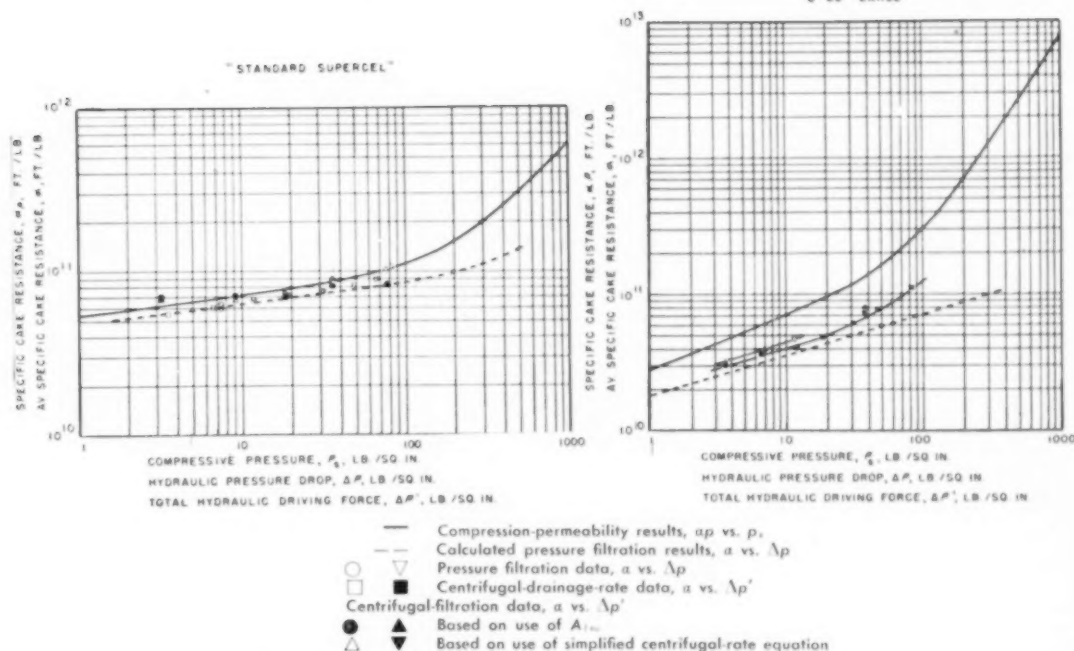
Fig. 6. Comparison of pressure and centrifugal-filtration data

extrapolated to higher compressive pressures for sizing the plant machine. The increase in compressive pressure resulting from increased weight of cake solids per unit area in plant machine must also be considered. One possible method of avoiding this extrapolation is operation of the small centrifuge at higher gravities than for the plant machine in order to approximate the total compressive pressure to be developed in the plant machine. The cake-holding capacity of the conventional 12-in. laboratory centrifuges, however, has limited the driving force which could be developed at 2000 g in such a machine to 100 lb./sq.in., which is much less than that possible in a plant-size machine operated at only 1,000 g. With the special centrifuge developed under this study, the cake-holding capacity has been increased so that driving forces up to 210 lb./sq.in. can be developed in a small machine, and such an approach now appears practical for sizing plant machines expected to operate to 1,000 g.

Although complete definition of centrifugal-filtration results from compression-permeability data has not in general been found possible, the compression-permeability data on the nature of cake properties greatly assist in defining whether materials of subsieve particle size are suited to centrifugal filtration. Thus materials yielding a value of a_p greater than 5×10^{10} ft./lb., at the maximum compressive pressure expected in the plant centrifuge, are probably not generally suitable to centrifugal

filtration. Similarly, definition of cake compressibility by means of compression-permeability measurements permits at least a semiquantitative prediction of the effect of centrifuge operation on expected performance and should immediately eliminate consideration of high-gravity operation with many materials.

Summary

The problems of predicting and correlating centrifugal-filtration results have been considered from an analytical, an experimental, and a practical point of view. In summary, the conclusions drawn from this work are:

Experimental data on centrifugal filtration were correlated satisfactorily by the simplified rate equation developed here. With semicompressible and compressible cakes, average specific cake resistance for the experimental conditions employed was found only slightly greater than for comparable pressure-filtration conditions.

Under certain centrifugal-filtration conditions, the simplified rate equation can be expected to fail with compressible materials, but the experimental conditions of this study did not cover the range necessary to define the point of failure.

Analysis of centrifugal-filtration conditions showed that, for the general case of centrifugal filtration, a cannot be calculated directly from compression-permeability data alone, as is possible for pressure-filtration conditions. For cer-

tain simplified cases this appears possible, but in general a knowledge of $f_s = f(r)$ is needed.

The field of application of centrifugal filtration with respect to subsieve-size materials was more clearly defined and the problem of centrifuge sizing with such materials was considered. Improved equipment and techniques for general solution of this sizing problem have been suggested.

Notation

A = cross-sectional area of cake perpendicular to direction of flow

A_c = active filter-medium area in perforate-basket centrifuge (cylindrical filtering surface of the bowl), sq.ft.

A_{lm} = log mean cross-sectional area of cake perpendicular to direction of flow, sq.ft.

A_m = arithmetic average cross-sectional area of cake perpendicular to direction of flow, sq.ft.

d = differential operator

f = function

g = acceleration in gravitational field, ft./(sec.)^2

g_c = conversion factor in Newton's law of motion, 32.2 (lb.)(ft.)/(lb. force)(sec.)^2

h = inside depth (parallel to axis) of cylindrical perforate-basket centrifuge bowl, ft.

K_c = average cake permeability of Burak & Storrow for centrifugal filtration, lb./sec.^{2} [$K_c = \rho_f g / a \rho_s (1 - \epsilon)$]

L = cake thickness in direction of flow, ft.

M = weight of filtrate collected after filtra-

Axis labels: CENTRIFUGAL FORCE AT BASKET WALL, F_o, LB FORCE / LB MASS (left), TOTAL HYDRAULIC DRIVING FORCE DEVELOPED, $\Delta P'$, LB /SQ IN (right)

CENTRIFUGE SPEED, N, REV./MIN.

———— 13-in. diam. laboratory centrifugal filter
$r_o = 6.5$ in.
$r_c = 2.5, 4.5,$ and 5.5 in.
— — — 40-in. diam. plant centrifugal filter
$r_o = 20$ in.
$r_c = 14, 16,$ and 18 in.
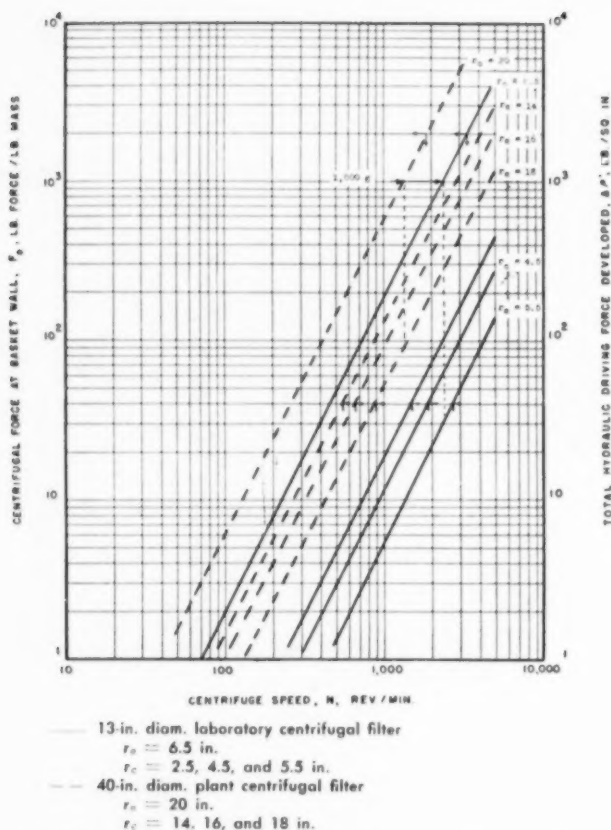
Fig. 7. Sizing effect on hydraulic driving force in centrifugal filtration (operation conditions: $r_f = r_c$)

tion time, θ, lb.

N = centrifuge speed, rev./sec.

p = hydraulic pressure at a point in cake, lb./sq.ft.

p_o = hydraulic pressure on discharge side of filter medium, lb./sq.ft.

p_1 = hydraulic pressure on cake side of filter medium, lb./sq.ft.

p_2 = hydraulic pressure at exposed surface of filter cake, lb./sq.ft.

p_g = hydraulic-pressure component developed by liquid flowing through cake, through action of centrifugal field, lb./sq.ft.

p_s = total mechanical compressive pressure developed in cake structure in direction of flow, lb./sq.ft.

p_v = hydraulic-pressure component developed by liquid flowing through cake, through a change in kinetic energy of liquid as flow through cake occurs, lb./sq.ft.

p_w = mechanical compressive pressure developed in cake structure, in direction of flow, resulting only from weight of cake solids acted upon by centrifugal field, lb./sq.ft.

Δp = total hydraulic-pressure drop across cake and filter medium in pressure filtration, lb./sq.ft.

Δp_c = hydraulic-pressure drop across cake in pressure filtration, lb./sq.ft.

Δp_m = hydraulic-pressure drop across filter medium in pressure or centrifugal filtration, lb./sq.ft.

$\Delta p'$ = total over-all hydraulic driving force causing flow through cake and filter medium in centrifugal filtration, lb./sq.ft.

$\Delta p'_c$ = over-all hydraulic driving force causing flow through cake in centrifugal filtration, lb./sq.ft.

Q = instantaneous rate of filtrate flow through cake cu.ft./sec.

Q_w = instantaneous weight rate of filtrate flow through cake, lb./sec.

r = radius to any point within cake when filtering on a cylindrical surface, ft.

r_o = radius to cake side of the filter medium surface when filtering on a cylindrical surface, ft.

r_c = radius to exposed cake surface when filtering on a cylindrical surface, ft.

r_f = radius to exposed surface of feed suspension or liquid when filtering in a perforate-basket centrifuge, ft.

R = over-all resistance in pressure filtration, 1/ft.

R' = over-all resistance in centrifugal filtration, 1/ft.²

R_m = filter-medium resistance, 1/ft.

u = superficial flow velocity through cake, ft./sec.

W = total weight of solid particles in cake, lb.

a = average specific cake resistance of actual filter cake (weight basis), ft./lb.

a_p = specific cake resistance at any point in cake, ft./lb.

ϵ = average cake porosity or void fraction, dimensionless

ϵ_p = porosity, or void fraction, at any point in cake, dimensionless

θ = time of filtration, sec.

ρ_f = density of suspending liquid or filtrate, lb./cu.ft.

ρ_s = density of solid particles, lb./cu.ft.

μ = filtrate viscosity, lb./ft.sec.

ω = angular velocity, radians/sec.

Consistent units, suitable for direct use in all equations of this paper, are given in this notation. More readily grasped units have been used in plotting several curves shown in figures of this paper.

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Appendix A.—Centrifugal-Filtration Theory

The filtration-rate equation as derived for a constant filtration area can be written as

$$Q = \frac{A g_c \Delta p}{\mu \left(\dfrac{aW}{A} \right) + R_m} = \text{cu.ft./sec.}$$
(1)

For the case of a constant-pressure filtration, the over-all pressure drop can be expressed as

$$\Delta p = \Delta p_{cake} + \Delta p_{cloth} = \text{constant}$$

or at any instant

$$(p_2 - p_o) = (p_2 - p_1) + (p_1 - p_o)$$
(2)

where p_o = discharge pressure and p_2 equals hydraulic pressure at surface of cake.

Also, rearranging (1) for constant-pressure filtration gives

$$\Delta p = \frac{Q \mu a W}{g_c A^2} + \frac{Q \mu R_m}{A g_c}$$
(3)

where at any instant

$$p_2 - p_1 = \frac{Q \mu a W}{A^2 g_c}$$
(4)

$$p_1 - p_o = \frac{Q\mu R_m}{Ag_c} \qquad (5)$$

and

$$\frac{p_1 - p_o}{p_2 - p_1} = \frac{AR_m}{a W} . \qquad (5a)$$

For any instant in a constant-pressure filtration cycle Equation (4) may be obtained by writing a pressure balance across the entire cake to obtain an over-all equation involving a. Equation (4) as written above is the balance for the case of a constant filter area, no work done on liquid flowing through cake and no change in kinetic energy of liquid flowing. Thus (4) is a special case of

$$p_1 + \frac{\rho_f u_1^2}{g_c} = p_2 + \frac{\rho_f u_1^2}{g_c} + p_g - p_f \qquad (6)$$

where in general

$$p_g = \frac{g}{g_c} \rho_f L. \qquad (7)$$

$$p_f = \frac{Q\mu a W}{A^2 g_c} \qquad (8)$$

Equation (6) simplifies to (4) when $u_1 = u_2$ and $p_g = 0$ as for the simple case of pressure filtration on vertical surface.

The problems under consideration here are twofold: first, the interpretation of compression-permeability data in terms of centrifugal filtration and, second, the development of a rate equation for centrifugal filtration which can be used to correlate actual centrifugal-filtration data in terms of a. In order to do this Equation (6) will have to be written to define a, the average specific cake resistance, in terms of centrifugal-filtration conditions since (4) will not apply.

For the case of a cylindrical perforate-

Fig. A. Perforate-basket centrifuge conditions.

basket centrifuge bowl as shown in Figure A, the area of the filter-medium surface, or the A term of Equation (5), is $A = 2\pi r_o h$, and at any instant

$$p_1 - p_o = \frac{Q\mu R_m}{2\pi r_o h g_c} \qquad (9)$$

since the thickness of the filter medium is negligible.

Rigorous analysis of the centrifugal-filtration case shows it to be a combination of Cases 2 and 3 of Part II of this paper, when $g/g_c \gg 1$. For the over-all cake one can write,

$$W = \pi h \rho_s (1 - \epsilon)(r_o^2 - r_c^2) \qquad (10)$$

and since g equals $\omega^2 r$, the centrifugal acceleration,

$$p_g = \frac{\rho_f \omega^2 (r_o^2 - r_c^2)}{2g_c} \qquad (11)$$

$$p_2 = \frac{\rho_f \omega^2 (r_c^2 - r_f^2)}{2g_c} . \qquad (12)$$

and

$$p_e = \frac{\rho_f (u_2^2 - u_1^2)}{2g_c} = \frac{\rho_f Q^2}{2g_c (2\pi h)^2}$$
$$\left(\frac{1}{r_c^2} - \frac{1}{r_o^2} \right). \qquad (13)$$

Also, substituting Equation (10) in (8) gives

$$p_f = \frac{Q\mu \pi h \rho_s (1 - \epsilon)(r_o^2 - r_c^2)}{g_c} \left(\frac{a}{A^2} \right) \qquad (14)$$

Writing the over-all pressure balance for the entire cake gives

$$(p_2 - p_1) = p_f - p_g - p_v. \qquad (15)$$

Substituting Equations (11) to (14) and rewriting gives

$$(p_2 - p_1) + \frac{\rho_f Q^2}{2g_c (2\pi h)^2} \left(\frac{1}{r_c^2} - \frac{1}{r_o^2} \right)$$
$$+ \frac{\rho_f \omega^2 (r_o^2 - r_c^2)}{2g_c}$$
$$= \frac{Q\mu \pi h \rho_s (1 - \epsilon)(r_o^2 - r_c^2)}{g_c} \left(\frac{a}{A^2} \right) \qquad (16)$$

which defines both a and A for this case.

Now considering an incremental cake thickness, dr, during any instant of a constant-pressure filtration, and writing the differential pressure balance, where a_p is the specific cake resistance in the increment, gives

$$-dp = -dp_g - dp_v + dp_f. \qquad (17)$$

For this cake increment, $A = 2\pi hr$, and $dW = \rho_s (1 - \epsilon_p) 2\pi hr\, dr$. Also,

$$u = \frac{Q}{2\pi hr}$$

$$du = -\frac{Q}{2\pi h} \frac{dr}{r^2}$$

and

$$dp_v = \frac{\rho_f u\, du}{g_c} = -\frac{\rho_f}{g_c} \left(\frac{Q}{2\pi h} \right)^2 \frac{dr}{r^3} \qquad (18)$$

$$dp_f = \frac{Q\mu a_p dW}{A^2 g_c} = \left(\frac{Q\mu \rho_s}{2\pi h g_c} \right)$$
$$a_p (1 - \epsilon_p) \frac{dr}{r} . \qquad (19)$$

Also, since $g = \omega^2 r$, and this centrifugal acceleration acts on the liquid in the direction of flow, then

$$dp_g = \frac{\rho_f \omega^2 r\, dr}{g_c} . \qquad (20)$$

Substituting Equations (18) to (20) in (17) gives

$$-\int_{p_1}^{p_2} dp = -\frac{\rho_f \omega^2}{g_c} \int_{r_o}^{r_c} r\, dr$$
$$+ \frac{\rho_f}{g_c} \left(\frac{Q}{2\pi h} \right)^2 \int_{r_o}^{r_c} \frac{dr}{r^3}$$
$$+ \left(\frac{Q\mu \rho_s}{2\pi h g_c} \right) \int_{r_o}^{r_c} a_p (1 - \epsilon_p) \frac{dr}{r} \qquad (21)$$

as the expression to be integrated, where p_1 is given by Equation (9) and p_2 is given by Equation (12) if it is assumed that liquid carried above cake surface is moving at basket speed. Integration is not possible without a knowledge of variation of a_p and ϵ_p with r. This is the basic rate equation for the general case of centrifugal filtration.

For the centrifugal-filtration case, $-dp \neq dp_s$ since compressive pressures other than the pressure drop act on the cake solids. In this instance the mechanical compressive pressure acting on the cake solids in the direction of flow results from both frictional drag dp_f and from a direct mechanical compressive pressure dp_w owing to the weight of cake solids suspended in liquid under the centrifugal field. Thus

$$dp_s = dp_w + dp_f. \qquad (22)$$

Substituting Equation (17) gives

$$dp_s = dp_w - dp + dp_g + dp_v. \qquad (23)$$

In these equations, dp_w results from centrifugal acceleration, $\omega^2 r$, acting on cake solids and is

$$dp_w = \frac{(\rho_s - \rho_f)(1 - \epsilon_p) \omega^2 r\, dr}{g_c} . \qquad (24)$$

Substituting Equations (18), (20), and (24) in (23) gives

$$-p_s = \int_{p_s}^{o} dp_s = \frac{(\rho_s - \rho_f)(1 - \epsilon_p) \omega^2}{g_c}$$
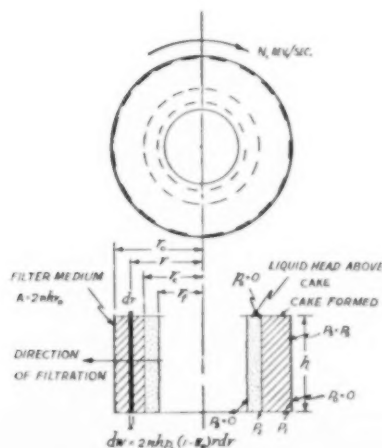
$$\int_{r_a}^{r_c} r\,dr - \int_{p_1}^{p_2} dp + \frac{\rho_f \omega^2}{g_c}$$

$$\int_{r_a}^{r_c} r\,dr - \frac{\rho_f}{g_c}\left(\frac{Q}{2\pi h}\right)^2 \int_{r_a}^{r_c} \frac{dr}{r^3} \quad (23)$$

to be integrated and solved for mechanical compressive pressure developed within cake at surface of filter medium, that at surface of filter cake being essentially zero.

Substituting Equations (19) and (24) in (22) and rearranging gives

$$\int_{p_s}^{o} \frac{dp_s}{(1-\epsilon_p)} = \frac{(\rho_s - \rho_f)\omega^2}{g_c} \int_{r_a}^{r_c}$$

$$r\,dr + \frac{Q\mu\rho_s}{2\pi h g_c} \int_{r_a}^{r_c} a_p \frac{dr}{r}. \quad (24)$$

Equation (23) cannot be integrated without a knowledge of $\epsilon_p = f(r)$, and even if the limit of p_s could be obtained from Equation (23), solution of Equation (24) from compression-permeability data would not be possible; a knowledge of $a_p = f(r)$ would still be required. Thus solution in the general case is not possible from compression-permeability data alone.

Solution from Compression - Permeability Data for Special Case. For the simplified case where $dp_w \ll dp_f$, as might result with low ρ_s solids or thin cakes and considerable liquid head carried above cake, Equation (24) simplifies to

$$\int_{p_s}^{o} \frac{dp_s}{(1-\epsilon_p)a_p} = \frac{Q\mu\rho_s}{2\pi h g_c} \int_{r_a}^{r} \frac{dr}{r}$$

$$= \frac{Q\mu\rho_s}{2\pi h g_c} \ln \frac{r_c}{r_a}. \quad (25)$$

Substituting Equation (16) then gives

$$\int_{p_s}^{o} \frac{dp_s}{(1-\epsilon)a_p}$$

$$= \frac{\ln \dfrac{r_c}{r_a}}{2(\pi h)^2(1-\epsilon)(r_a^2 - r_c^2)}\left(\frac{A^2}{a}\right)$$

$$\left[(p_2 - p_1) + \frac{\rho_f Q^2}{2g_c(2\pi h)^2}\left(\frac{1}{r_c^2} - \frac{1}{r_a^2}\right)\right.$$

$$\left. + \frac{\rho_f \omega^2(r_a^2 - r_c^2)}{2g_c}\right]. \quad (26)$$

If r_a is relatively large compared to $r_a - r_c$, then the kinetic-energy-change term of Equations (23) and (26) becomes very small. Also Equation (26) can be rearranged in form:

$$\int_{o}^{p_s} \frac{dp_s}{(1-\epsilon_p)a_p} = \left[\frac{A^2}{(A_{lm})(A_m)}\right]$$

$$\left[\frac{(p_2 - p_1) + \dfrac{\rho_f \omega^2(r_a^2 - r_c^2)}{2g_c}}{(1-\epsilon)a}\right]. \quad (27)$$

Then if undefined A^2 term of Equation (16) is taken as $A_{lm} \times A_m$, Equation (27) simplifies to

$$\int_{o}^{p_s} \frac{dp_s}{(1-\epsilon_p)a_p}$$

$$= \frac{(p_2 - p_1) + \dfrac{\rho_f \omega^2(r_a^2 - r_c^2)}{2g_c}}{(1-\epsilon)a}. \quad (28)$$

This can be integrated and solved numerically from compression-permeability data since for this special case, Equation (23) simplifies to,

$$p_s = - \int_{p_1}^{p_2} dp + \frac{\rho_f \omega^2}{g_c} \int_{r_a}^{r_c}$$

$$r\,dr = (p_2 - p_1) + \frac{\rho_f \omega^2(r_a^2 - r_c^2)}{2g_c} \quad (29)$$

and this can be solved directly. For this special centrifugal-filtration case the total hydraulic driving force across the cake, $\Delta p_c'$, equals the maximum mechanical compressive pressure developed in cake and

$$\Delta p_c' = p_s = (p_2 - p_1)$$

$$+ \frac{\rho_f \omega^2(r_a^2 - r_c^2)}{2g_c}. \quad (30)$$

Substituting this in Equation (28) gives

$$\int_{o}^{\Delta p_c'} \frac{dp_s}{(1-\epsilon_p)a_p} = \frac{\Delta p_c'}{(1-\epsilon)a} \quad (31)$$

which is easily solved from compression-permeability data.

$$Q = \frac{\Delta p' g_c}{\left[\mu\left[\dfrac{2\pi h(r_a - r_c)}{\ln \dfrac{r_a}{r_c}}\right]\dfrac{2\pi h(r_a - r_c)}{2}\right] + \dfrac{R_m}{2\pi h r_a}}$$

Simplied Rate Equation for Special Case of Nearly Incompressible Cake. Although the general rate Equation (21) can not be solved without a knowledge of $a_p = f(r)$ and $\epsilon_p = f(r)$, solution for the special case of a nearly incompressible cake is possible, since for this case $a_p \approx a$ and $\epsilon_p \approx \epsilon$ at any point in cake. Thus Equation (21) can be integrated directly where Equations (9) and (12) define p_2 and p_1 if $p_a = 0$ is assumed. This

gives

$$\frac{Q\mu R_m}{2\pi r_a h g_c} - \frac{\rho_f \omega^2(r_c^2 - r_f^2)}{2g_c}$$

$$= \frac{\rho_f \omega^2(r_a^2 - r_c^2)}{2g_c} - \frac{\rho_f}{2g_c}\left(\frac{Q}{2\pi h}\right)^2$$

$$\left(\frac{1}{r_c^2} - \frac{1}{r_a^2}\right) - \frac{Q\mu a\rho_s(1-\epsilon)}{2\pi h g_c} \ln \frac{r_a}{r_c}. \quad (32)$$

Rearranging gives

$$Q =$$

$$\frac{\pi h \rho_f \omega^2(r_a^2 - r_f^2) - \dfrac{\rho_f Q^2}{4\pi h}\left(\dfrac{1}{r_c^2} - \dfrac{1}{r_a^2}\right)}{\mu\left[a\rho_s(1-\epsilon)\ln \dfrac{r_a}{r_c} + \dfrac{R_m}{r_a}\right]}. \quad (33)$$

Since for even the special laboratory centrifuge the kinetic-energy-change term in the numerator is very small, it can be neglected. Noting this and that $\omega = 2\pi N$, one can write the final rate equation for this simplified case as

$$Q = \frac{4\pi^3 N^2 \rho_f h(r_a^2 - r_f^2)}{\mu\left[a\rho_s(1-\epsilon)\ln \dfrac{r_a}{r_c} + \dfrac{R_m}{r_a}\right]}. \quad (34)$$

In order to compare this simplified centrifugal-filtration-rate equation with the standard filtration-rate equation (Equation (1)), one can consider the total hydraulic driving force, $\Delta p'$, to replace Δp, where

$$\Delta p' = \frac{\rho_f \omega^2(r_a^2 - r_c^2)}{2g_c} + \frac{\rho_f \omega^2(r_c^2 - r_f^2)}{2g_c}$$

$$= \frac{\rho_f \omega^2(r_a^2 - r_f^2)}{2g_c}. \quad (35)$$

Noting from Equation (10) that

$$\rho_s(1-\epsilon) = \frac{W}{\pi h(r_a^2 - r_c^2)} \quad (36)$$

and substituting Equations (34) and (35) in (33) gives

which can be written as

$$Q = \frac{\Delta p' g_c}{\mu\left(\dfrac{aW}{A_{lm} \cdot A_m} + \dfrac{R_m}{A_c}\right)}$$

Thus it appears that the A^2 factor entering into the cake-resistance term should be made up of $A_{lm} \times A_m$ while that entering the cloth-resistance term should be A_c, at least for the special case of centrifugal filtration for which Equation (33) was developed.

MIXED-BED ION EXCHANGE

A Symposium Covering Theory, Design of Production Units, and Production Results

Joseph Thompson, Francis X. McGarvey, John F. Wantz, S. F. Alling, M. E. Gilwood, and David R. Babb

Part II

Production Results

REGENERATION TECHNIQUES *

The chief factors involved in the selection of a mixed-bed water-conditioning unit are the composition of the influent solution, the desired characteristics of the effluent, and the cost of the treatment. Economic considerations may dictate removal of alkalinity before the mixed-bed unit is used for complete ion removal. This can be accomplished by passing all or a portion of the water through a cation exchanger operating in the hydrogen cycle and removing carbon dioxide by aeration.

In addition to selection of a system of treatment, composition of water has an effect on the performance of the beds and will influence the design of the mixed bed itself. This is particularly important when the procedure of regeneration is established. Various methods have been suggested to eliminate the possibility of precipitation of calcium carbonate, calcium sulfate, and magnesium hydroxide during regeneration (5).

Rinse procedures are similar regardless of the method of regeneration, but appreciable savings in rinse-water requirements are possible with the mixed-bed technique, because the initial rinse is terminated before the last traces of excess regenerant are washed from the cation- and anion-exchange resins. At this point the resins are mixed, and the final rinse actually promotes the removal of the remaining regenerant by ion exchange. This technique may reduce rinse requirements by as much as 50%.

The characteristics of the influent solution have only a limited effect on the completeness of ion removal during the exhaustion of a mixed-bed unit. Under normal operating conditions, quality of effluent may be considered independent of influent-solution composition and will be dependent on type of resins used in the system. This is not true of multiple-bed exchangers where effluent qualities are closely related to influent characteristics, regardless of the type of resins used.

* From "Theory of Monobed Exchange," by Joseph Thompson and F. X. McGarvey.

TYPICAL OPERATION

Strong-acid—strong-base Exchange Resins. This combination has been used extensively in water conditioning, for it economically reduces electrolytes and silica below values attainable by any other procedure presently available for the production of water on a commercial scale.

Silica leakage from multiple-bed strong-base exchangers is influenced greatly by the pH of the water passing through the bed. This in turn is affected by the leakage of ions from the cation-exchange-resin bed. The mixed-bed technique eliminates this difficulty by permitting only very minor pH variations in the bed. It is not surprising that mixed beds of this type should give much lower residual silica concentrations than are usually found with multiple-bed systems. Table 7 shows the effect of

TABLE 7.—EFFECT OF CATION LEAKAGE ON SILICA LEAKAGE FROM A MULTIPLE-BED SILICA-REMOVING DEIONIZING SYSTEM

Cation leakage, p.p.m. as CaCO₃	Final pH of deionized effluent	Silica leakage, p.p.m. as SiO₂*	
		IRA-400	IRA-410
1	7.7	0.01	0.04
2	8.2	0.02	0.08
3	8.6	0.02	0.13
4	9.0	0.03	0.17
5	9.4	0.04	0.25
6	9.6	0.06	0.30
7	9.8	0.10	0.38
8	10.0	0.15	0.45
10	10.3	0.20	0.65
12	10.5	0.30	1.50
16	10.7	0.40	high
18	11.0	1.00	high

* These values are obtainable with only very large excesses of regenerant.

cation leakage on silica leakage from multiple-bed silica-removing deionization systems.

Joseph Thompson and Francis X. McGarvey are with Rohm & Haas Co., Philadelphia, Pa.; John F. Wantz is with Illinois Water Treatment Co., Rockford, Ill.; S. F. Alling with Hungerford and Terry, Inc., Clayton, N. J.; M. E. Gilwood with The Permutit Co., New York; and David R. Babb with Elgin Softener Corp., Elgin, Ill.

The data indicate that for completeness of silica removal at practical regeneration levels an effluent pH of 7.0 to 7.5 is necessary. The most convenient way to maintain these optimum conditions is by mixed-bed deionization. Typical curves for a mixed-bed unit utilizing resin IR-120 and IRA-410 are shown in Figure 9. The treated water is characterized by very low residual silica, i.e., less than 0.01 p.p.m. with the pH controlled at 7.0. A multiple-bed system under the same conditions would not reduce the silica to less than 0.05 to 0.1 p.p.m.

Strong-base—weak-acid Exchange Resins. In certain cases a mixture of strong-base and weak-acid exchangers has particular advantage. This is true for deionization of sugar solutions, since the carboxylic-type exchanger does not cause inversion (1). When used for the treatment of waters, a specific resistance of about 5×10^5 to 1×10^6 ohms/cm. has been recorded. The carboxylic exchanger has high regeneration efficiency with acid but is slightly rate sensitive owing to slow rates of diffusion encountered. For good operation, the cation resin must be fully regenerated.

Weak-base—strong-acid Exchange Resins. This system has many interesting possibilities in water conditioning, since the anion exchanger may be regenerated at a high efficiency with sodium hydroxide, sodium carbonate, or ammonium hydroxide. Since this type of mixed bed does not remove weak acids or silica, it is economical for process-water treatment where extreme quality is not necessary. A typical curve for this system is shown in Figure 10. The resin employed will pick up carbon dioxide at the initial portion of the run, but the weak acid will be pushed off as the run continues, and an enrichment of the effluent in carbon dioxide will occur.

Weak-acid—weak-base Exchange Resins. Considerable interest has been expressed in a mixture of weak-acid and weak-base exchangers, since this combination will give almost 100% efficiency in regeneration. As these materials have very slow rates, small particle sizes, long contact
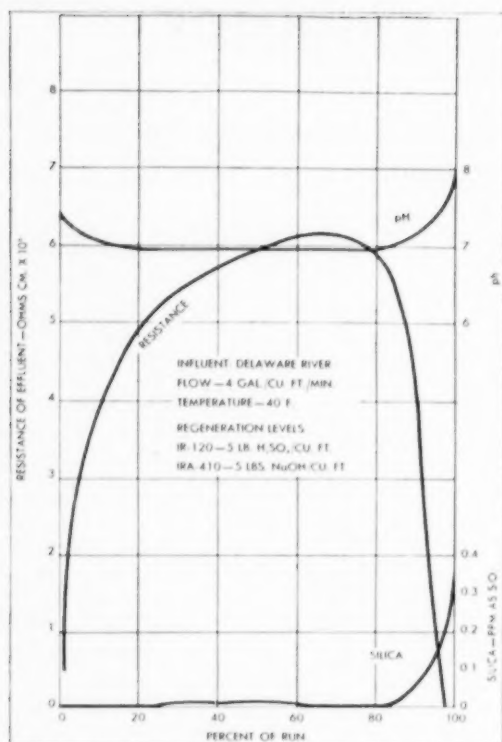
Fig. 9. Performance of strong-acid—strong-base ion-exchange resins in mixed-bed unit.
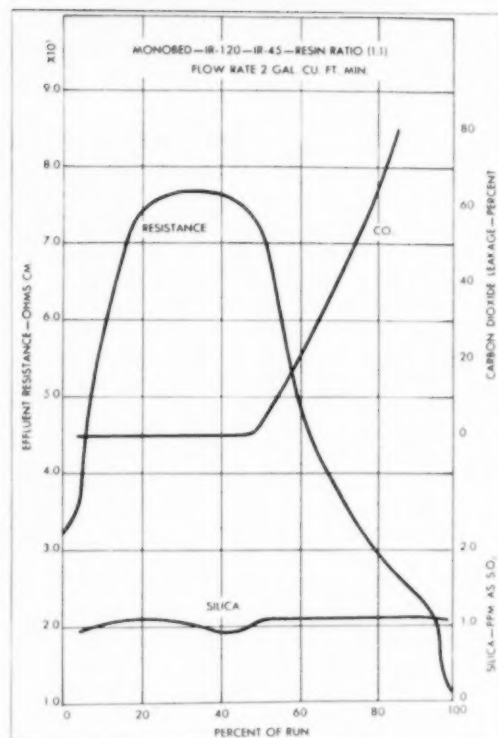
Fig. 10. Performance of strong-acid—weak-base ion-exchange resins in mixed-bed unit.

time, and high concentrations are needed for efficient deionization. This system is particularly useful if deionization without pH changes is desired.

INFLUENT QUALITY *

In the past too little emphasis has been placed on the need of using good-quality, clear, pure water through mixed-bed demineralizer units. Resins on the market today are well made and reasonably rugged, but they are made purely for ion-exchange work and not for filtering water. The influent water going to a mixed-bed unit should have a turbidity of less than 5 p.p.m., no oil, no bacteria of the crenothrix type, less than 0.3 p.p.m. residual alumina (Al_2O_3), and less than 0.5 p.pm. residual chlorine. If such a water is not available, prefilters with or without pretreatment should be furnished as a protection for the costly resins supplied with the unit.

If the amount of residual chlorine in the incoming water is high, for example over 0.5 p.p.m. average, then an activated-carbon filter or pretreatment with some kind of antichlor should be provided. This figure of 0.5 p.p.m. may be considered low by many, but until the exact tolerance of anion-exchange resins

* From "Engineering Results with Mixed-Bed Exchanger," by S. F. Alling.

for chlorine has been definitely established, it is better to be safe, especially in view of the small additional cost of a prefilter or pretreatment with an antichlor.

If the mixed-bed unit is small and the incoming water clear, but the amount of residual chlorine somewhat dangerous, the feeding of a trace of sodium thiosulfate is simple, effective, and, in the overall picture, reasonably economical. At installations where the water also contains some suspended matter as well as chlorine, a carbon filter that will clarify the water and at the same time dechlorinate it is to be preferred. If there are bacteria present, as for example crenothrix, just sufficient chlorine and alum should be applied to destroy the bacteria and coagulate the finely divided cell structure. The water should then be passed through a sand filter to complete the clarification process before it is demineralized.

One of the most troublesome impurities in water passing through a resin bed is residual alumina, and, since alum is used at many municipal water-filtration plants as well as at private water systems, some residual alumina is frequently found in the water supplies used through mixed-bed units.

Pitfalls. Several years ago two large sodium-cycle resin units were supplied

for use on water that apparently was clear, had a normal temperature, contained no chlorine, and seemed ideal for use through new equipment. Operation was perfect, all design factors were standard, and still the capacity of the resin dropped nearly 50% in less than 6 months of operation. Acid treatment of the resin was completely ineffective, but when the influent water was tested constantly for about 4 hours, from 0.3 to 0.5 p.p.m. residual alumina was found to be present in the water going to the resin units. The water at its source was well water that had been aerated and treated with lime and then a little filter alum for coagulation purposes. Some of the alum was not completely precipitated at the high operating pH and was coming through the filters in solution or as a colloid. This residual alumina was surface coating, or casehardening, the resin and reducing its receptiveness to the regenerant and consequently its exchange capacity in exactly the same way that residual alumina will increase the physical hardness and reduce the exchange capacity of natural greensand zeolites.

In a mixed-bed unit, the acid and alkali regenerants tend to keep the resins clean and free of the harmful effect of residual alumina, but the presence of more than 0.3 p.p.m. residual alumina is

harmful, from a performance standpoint, to the resins. Water to be demineralized through a mixed-bed unit should be first clarified of any residual alumina present.

Similar experience has been encountered with turbidity, oil, iron, and oil-bearing bacteria in water, and in connection with the latter, or with any form of bacterial or algal plant life, a microscopic examination is recommended for water that is to be demineralized with a mixed-bed unit. Inorganic iron in water is not troublesome, but crenothrix, which requires only 0.3 p.p.m. of iron for its sustenance, is a serious factor. It is important, therefore, to make a careful analysis, both chemical and microscopic, of each sample of water, before a unit containing any of the present-day synthetic ion-exchange resins is recommended. In the case of special problems where operational data are not available, a pilot unit should be set up, the type of water that is to be treated synthetized, and a series of operating runs conducted. On the basis of these results effluent guarantees and performance data can be easily and accurately estimated.

EFFLUENT QUALITY

Table 8 shows some of the test results obtained from a 5-ft.-diam. by approximately 10-ft.-high mixed-bed unit.

This particular mixed-bed demineralizer unit contained 40 cu.ft. of a nuclear sulfonic-type cation-exchange resin and 43 cu.ft. of a strongly basic amine-type anion resin, and at a flow rate of 50 gal./min. the effluent showed a specific resistance of about 4,000,000 ohms, while at a flow rate of 83 gal./min., the resistivity in the treated water was over 16,000,000 ohms. Further tests are being conducted on the same unit to determine the effluent quality at all flow rates from the very smallest to the maximum, but preliminary data obtained to date indicate that it is better to operate these mixed-bed units at a medium-to-high rate of flow than at a very low rate.

Fig. 11. Test run in series employing each process.

Should this be the case, it is going to be possible to supply industry with smaller size, lower cost units to accommodate the flow-rate requirements.

In the tests referred to here on the 5-ft.-diam. mixed-bed test unit, all specific-resistance readings were determined on a special conductivity bridge. With the conventional type of solu-bridge, the specific-conductance reading was consistently 0 μ mhos. With an inlet-water silica content of 8.0 p.p.m., the silica in the effluent by spectrophotometer tests was always less than 0.02 p.p.m., while the 30 p.p.m. of free carbon dioxide in the inlet water was reduced to 0 in passing through the mixed-bed unit.

Regenerant dosages and resin-capacity figures recommended by the resin manufacturer in both the laboratory and pilot-plant units were met with a liberal factor of safety in these test runs.

DEMINERALIZING FOR HIGH-PRESSURE BOILER FEED WATER *

Experience and research in operating boilers at pressures above 1,000 lb./sq.in. have indicated the importance of maintaining silica concentration in boiler saline at levels below 5 p.p.m. (8) Higher silica levels result in solution of silica in the steam and subsequent silica deposits on the turbine blades in the lower pressure stages of the turbines. At pressures of 1,800 to 2,000 lb./sq.in. and higher, the maximum silica permitted in the saline may frequently be less than 1 p.p.m. Since the make-up water is concentrated many times in the boiler, engineers frequently specify that the treated

* From "Mixed-Bed Demineralizing of High-Pressure Boiler Feed Water," by M. E. Gilwood.

make-up water must contain no more than 0.1 p.p.m. silica.

The mixed-bed demineralizing process, employing a mixture of cation and anion exchangers in a single bed, is of special interest because it provides in effect a series treatment by a great number of individual demineralizing steps, with the result that electrolyte impurities are reduced to the vanishing point. The purified water thus contains so few inorganic impurities that they cannot be determined accurately by conventional water-analysis methods but are checked only by conductivity measurements. Since highly basic-anion exchangers can be employed in the mixed-bed demineralizer, the silica content of a water can also be reduced to extremely low levels. The commercial development of mixed-bed demineralization has been furthered (1) in recent years.

The power-plant engineer does not attach much importance to the ability of a demineralizer to produce water of extremely low conductivity or high resistivity. In fact, he most often adds some chemicals, such as phosphates, sulfites, and hydroxide, to protect the boiler against corrosion and deposits. Modern demineralizing is important in this field because silica can be removed from water to levels of 0.1 p.p.m. dissolved or lower.

A fully automatic mixed-bed demineralizer was recently installed at a large steam electric station in central New York State. Two mixed-bed units, 42-in. diam. by 7-ft. high are provided, each capable of treating 48 gal./min. of filtered raw water. This plant is to provide all the make-up water for a 1,450 lb./sq.in. boiler having about 0.5% make-up.

TABLE 8.—TEST RESULTS
(p.p.m. as $CaCO_3$)

	Influent	Effluent
Ca	14.0	0
Mg	16.0	0
Total hardness	30.0	0
Na	27.0	0
HCO_3	7.0	0
SO_4	20.0	0
Cl	30.0	0
CO_2 (as CO_2)	30.0	0
SiO_2	9.0	0.02
pH	5.9	7.0
Spec. resistance (ohms/cm.)		Over 16,000,000
Conductivity (μ ohms)		0.06

Considerable test work has been done to determine how effectively silica can be reduced by mixed-bed demineralization. With careful operation and warm regeneration, it is possible to reduce the silica content of the water to as low as 0.01 p.p.m. dissolved silica. Silica in these low amounts must be measured by the modified Bunting colorimetric technique described by Robison, Pirsh, and Grimm (6) employing a spectrophotometer operating at 820 mμ. Even then, there is considerable doubt as to the accuracy of the silica measurements in this range.

Parallel tests have been made treating the same water by the same anion-exchange resin * in a mixed-bed demineralizer and a two-step demineralizer. In these tests some slight variation in effluent silica was found to occur on successive runs. Figure 11 illustrates one of the best test runs obtained in a series of runs employing each process. In this series of runs, the effluent silica varied from 0.005 to 0.02 p.p.m. by each method. The resistivity of the mixed-bed demineralized water varied between 6 and 13 megohms/cm., whereas the resistivity of the two-step demineralized water varied from 0.25 to 0.5 megohm/cm. during the runs.

Influent water:

MO alkalinity	$=$	118 p.p.m. as $CaCO_3$
Carbon dioxide	$=$	2 p.p.m. as CO_2
Silica	$=$	16 p.p.m. as SiO_2
Chloride + sulfate	$=$	14 p.p.m. as $CaCO_3$

Anion regenerant: 6 lb. sodium hydroxide/ cu.ft. applied as 5% solution at 110° F.

These test results indicate that it is possible to obtain demineralized water containing as low as 0.01 p.p.m. silica by mixed-bed demineralizing and also by two-step demineralizing in the best runs. It undoubtedly requires extremely careful operating control always to obtain results approaching these in continuous large-scale plant operation.

It is not difficult to obtain such extremely low silica residuals by either mixed-bed or two-step demineralization when waters containing a comparatively low silica/total anions ratio, i.e., less than 1/10 to 1/5, is treated. Where the silica/total anion ratio is much higher, it becomes more difficult to maintain the residual silica at the 0.01 p.p.m. levels, and more vigorous regeneration conditions must be maintained.

In mixed-bed demineralizing, the bicarbonates are not usually removed from the water prior to contact with the anion exchanger, as they are in two-step demineralizing. The silica/total anion ratio at the anion exchanger when the same water is treated by the two methods, will be much lower with mixed-bed demineralization. Thus, low-silica residuals should be obtained more readily in mixed-bed demineralizing. However, this advantage of mixed-bed demineralizing can be obtained only as a result of higher chemical operating cost; i.e., sodium hydroxide is needed for bicarbonate and carbon dioxide in influent water. It appears that similar ease in removing silica and obtaining low silica levels is attained in both processes if the loading of silica on the anion exchanger is kept low in the two-step process. This can be done by cutting off the anion-exchange runs of the two-step treatment somewhat before the normal exhaustion point.

Figures 12 and 13 illustrate a comparison of mixed-bed– and two-step–demineralizing chemical operating and installation costs on various waters, listed in Table 9. It is evident that the equipment cost for mixed-bed demineralizing is lower than for two-step demineralizing when a low-solids water like the New York City Catskill supply is treated. This is balanced to some extent by increased chemical-operating cost. Nevertheless, it appears that a lower over-all cost can be obtained by mixed-bed demineralizing for such a low-solids water. This study also indicates that when other typical water supplies are treated, mixed-bed demineralizing is somewhat more costly than two-step demineralizing both in first cost and in chemical-operating cost.

Mixed-bed demineralizing thus seems to be of advantage for producing low-silica – low-electrolyte high-pressure boiler feed water where the raw-water solids are low. Further tests and large plant experience are required to determine whether it is possible continuously to produce water containing less than 0.01 p.p.m. silica by mixed-bed demineralizing.
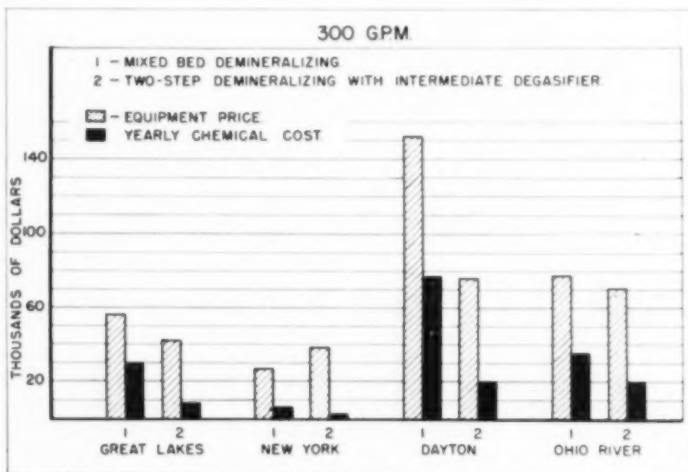
* In this case Permutit S-1, a trade name of The Permutit Co.

Fig. 12. Cost comparison between mixed bed and 2-step with degasifier.

Fig. 13. Cost comparison between mixed bed and 2-step with degasifier.

APPLICATION *

Aside from medical, pharmaceutical, and specialized chemical-process applications for purification of organic and inorganic mixtures and compounds, the primary use of mixed-bed ion exchange has been in the field of water purification. Applications in specialized fields for the most part make use of the same marked advantages that have made the mixed bed advantageous for water treating, plus the outstanding feature that process liquid passes through the bed at a relatively neutral pH.

Conventional ion-exchange demineralization employing first a cation-exchange column, followed by an anion-exchange column, has certain shortcomings that were recognized in the industry simultaneously with the development of industrial deionization equipment. The cumbersome four-tank deionizers that incorporated additional scavaging beds to remove ionic-leakage products were

tration in the influent to a multiple-tank deionizer is too high, pronounced leakage will occur which results in low-quality deionized effluent.

As a result of these shortcomings, it will be seen that the use of multiple-tank deionization is disadvantageous, from at least an operating standpoint, on both very high-calcium waters and soft waters containing an appreciable per cent of sodium ions.

Mixed-bed deionization on the other hand injects its characteristic advantage of eliminating leakage due to the multiple-layer scavenging effect similar in nature to the theoretical plates of distillation theory and lends itself readily to the use of soft-water influent to prevent the precipitation of calcium salts during regeneration. Operating experience has shown that best possible results are achieved when mixed-bed deionizers are operated on zeolite-softened influent whenever an appreciable per cent of calcium is present in the raw influent water.

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Discussion

Anonymous: I note that the symposium covers ion exchange, but the discussion has been pretty much limited to the treatment of water. I am particularly interested in the application of ion-exchange resins to the treatment of organic materials with reference to removing organic or inorganic impurities from organic liquids, for example, iron from the monomer, or perhaps an allied organic material from the organic compound.

Joseph Thompson (Rohm & Haas, Philadelphia, Pa.): I'd like to mention why most of the discussion on commercial applications was applied to the treatment of water. Actually, the treatment of water is the least complex of monobed-deionization-system applications. This is particularly important when you consider that the various equipment-manufacturing organizations had to fix their engineering design. Now that the engineering design has been fixed and units are commercially available and the bugs have been ironed out of the operation, I think you'll see the possibility for applications to the treatment of solutions other than water.

On the problem of the removal of ions from organic materials, there is some question in my mind as to whether you are referring to a straight organic liquid, whether you actually have some moisture present, whether you have materials in an ionized form, or what the conditions of operation are. Heretofore we have thought only of removal of ions from aqueous media, but, as we are becoming more familiar with these ion-exchange resins and their applications, we find there are real possibilities in nonaqueous media. For example, it is possible to remove iron from benzene. It is possible to remove other contaminating constituents from organic liquids. Whether the monobed deionization system can be applied to your particular problem, I can't say positively. However there certainly exists the possibility of applications for this particular technique far beyond that of straight water conditioning.

John F. Wantz (Illinois Water Treatment Co., Rockford, Ill.): I think one of the greatest advances insofar as the treatment of organics is concerned has come about in the treatment of glycerine. Our firm has done a considerable amount of work in that line, and we found that the mixed-bed-type ion-exchange equipment has made possible the substitution of ion exchange for distillation processing of crude glycerine and the obtaining of a C.P. product. Undoubtedly the same process can be applied to other organics in solution.

TABLE 9.—TYPICAL WATER ANALYSES

	p.p.m. as	Great Lakes	New York (Catskill)	Dayton, Ohio	Ohio River
Total hardness	CaCO₃	124	20	370	126
Sodium	CaCO₃	16	5	20	72
M O alkalinity	CaCO₃	98	12	278	45
Chloride and sulfate (ThMA)	CaCO₃	42	13	112	153
Silica	SiO₂	4	2.4	10	16

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the result. Even now with the advent of more strongly basic anion exchangers and highly efficient cation exchangers, leakage and precipitation present a paradoxical situation that is difficult to overcome with conventional multiple-bed techniques.

For most industrial applications, it is desirable to employ low-cost sulfuric acid as a cation regenerant since the use of hydrochloric acid would entail both higher original equipment cost and higher operating costs. Calcium eluted from the cation resin during regeneration, however, tends to precipitate when sulfuric acid is used on normal raw waters. Although complicated regeneration techniques have been developed which tend to overcome precipitation difficulties, their use has been limited by inherent partial effectiveness. Even with the most exacting laboratory-controlled techniques, only incomplete regeneration of a calcium-laden bed may be obtained using economic quantities of sulfuric acid.

On the other horn of the dilemma, there is the problem of sodium-ion leakage. If per cent of sodium-ion concen-

The combined advantages of softening and simultaneous filtration not only prevent the precipitation of hardness compounds, but also prevent the deposition of suspended matter within the deionizer bed.

For these reasons, it is not surprising that mixed-bed deionization is being adopted in such common applications as high-purity water for chemical processing; resin polymerization; rinsing of organic chemical products; and in the manufacturing fields where rinsing, washing and processing of lamps, glass, and metal parts are required. High- and low-pressure boiler make-up, recirculating systems of all kinds, and chemical laboratories are making use of the mixed bed.

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A glycerine solution is treated in the same way as a water solution. The ionizable solids are removed along with the removal of color. Special preparation has to take place in preparing the solution for treatment. Removal of fatty acids is essential. In addition, it is necessary to have a clear turbidity-free solution.

M. E. Gilwood (The Permutit Co., New York, N. Y.): As I understand the question, I believe the query was about the use of ion exchange for removing the impurities from monomers Mixed-bed demineralizing has been used to a very small extent on truly nonaqueous media. The comment by Mr. Wantz concerning glycerine of course applies to a solution of a nonaqueous material in water. One can demineralize sugar solutions or glycerine solutions, etc.

Regarding purification of a monomer, however, we did do some work in connection with the removal of inhibitor in a styrene monomer solution, and it can be done by use of proper anion-exchange media. This has nothing to do with mixed-bed demineralization, the topic of this symposium. Mixed-bed demineralizing has not been applied to any great extent to applications other than in treatment and has not been applied to monomer purification. However, there has been some work done on the purification of nonaqueous media by specific ion exchangers in a single- or in a two- or three-step treatment.

I think everyone is familiar with the purification of formaldehyde by anion exchange. You might call that a monomer. Formic acid is removed from the formaldehyde solution to purify formaldehyde. Again, this is not a true nonaqueous system but really a solution of formaldehyde and water.

In connection with the purification of glycerine and similarly in the purification of other polyhydric alcohols like sorbitol, most of the steps in the treatment are carried out not by mixed-bed demineralization but two- or four-step demineralization and merely employ the mixed-bed demineralization as a polishing treatment at the end—to scour out the last traces of contaminating material to bring the product within specifications. The bulk of the purification work can be done in the more easily operable and maintainable single-step treatments where there is just one ion exchanger in a tank and the flow passes through successive tanks of the ion exchangers.

Horace Hinds, Jr. (Corn Products Refining Corp., Argo, Ill.): I notice that Mr. Wantz in discussing the design of his monobed exchanger suggested two windows. I wonder what the operator sees in those windows and what action is implied?

Wantz: Actually, I don't believe that from the standpoint of operation the windows are imperative. They do act as an aid, however, to the operator, particularly the sight glass at the interface. The operator can visually check to make sure that his backwash procedure has properly classified the ion-exchange materials. Insofar as the window or sight glass at the top of the bed is concerned, it helps the operator maintain a proper bed level. He can readily see whether or not his influent is clear and whether he might be collecting suspended solids on the top of the bed. It gives him a visual check inside the tank, something that has not been used on most conventional two-bed deionizers.

H. B. Gustafson (Infilco, Inc., Tucson, Ariz.): In the first place I want to thank Mr. Gilwood for bringing out the fact that the mixed-bed process is an addition to processes of water treatment and that it does not necessarily make obsolete or replace older methods. Choice of method and equipment should be made only after careful study of cost of treatment, cost of equipment, amount of water being treated, quality requirements, etc. Only after all these have been considered, can one determine whether conventional cation anion, mixed bed, or a combination is the best process.

Next I'd like the panel to consider this 5 p.p.m. of turbidity which Mr. Alling, I believe, suggested as standard. To me it seems a little bit incongruous to suggest a 5 p.p.m. turbidity standard for a water having a specific resistance of 17 million ohms. A turbidity of 5 p.p.m. represents about 500 times as much impurity, present as suspended matter, as the dissolved impurity contained in water having a specific resistance of 17 million ohms.

S. F. Alling (Hungerford & Terry, Inc., Clayton, N. J.): With respect to the question of turbidity, first of all, if your plant requires water with a specific resistance of anything like 10, 15, or 20 million ohms, you should have an influent water of 0.0 turbidity, by the Baylis test. On the other hand, there are many requirements where one might be using these mixed-bed units that do not require a water of anything like that quality, and in such cases I expressed the opinion that 5 p.p.m. maximum turbidity might be a figure to use. We have quite a number of these mixed-bed units in use where the quality of the water does not have to be anything like what we discussed to a very large extent. In those cases it would be foolish to put in pretreating equipment to take the turbidity in the influent water down to 0.0.

Gustafson: Mr. Schulze, have you ever had any problems in connection with residual tastes and odors coming from the resin exchangers used in mixed-bed units?

R. E. Schulze (Culligan Zeolite Co., Northbrook, Ill.): In potable waters, tastes and odors and colors are highly undesirable, and it was mentioned before that some of these nonionizable constituents may be or must be treated in some way other than by deionization at possibly a little more or a little less expense. However, we have in mixed-bed systems taken the taste problem connected with iron and completely eliminated it with mixed-bed deionization. The matter of color is pretty difficult to control with deionization; it would be much easier to control with activated carbon as a polishing absorbent filter either before or after the deionization process.

In connection with some of the other odors or tastes, no doubt you are familiar with the sulfide waters in the area north of Chicago, where there are waters with as high as 60 to 80 p.p.m. hydrogen sulfide. For control of odors or tastes at that particular level, it is usually found preferable to oxidize the hydrogen sulfide with chlorine to the sulfur state and then to filter it out with activated carbon, rather than to try to control it through deionization.

E. K. Erickson (DuPont Co., Martinsville, Va.): In the manufacture of nylon, Mr. Alling, we are well aware of the difficulties experienced by allowing bacteria to get into the process lines. I think you made the point that it was highly undesirable to allow great concentrations of bacteria to get into demineralizing equipment. I wonder if you could elaborate on that.

Alling: If there are active bacteria or algae in the influent water, there is a strong possibility of their lodging in cracks and crevices in the demineralizer unit, and eventually, if they are allowed to remain there, there is, in spite of a backwashing operation, a possibility of large growths that may plug the interior of the unit. In addition, there is always the possibility of the bacteria or algae coating over and fouling the resin grains, thus lowering their efficiency. True, regenerants will tend to keep the resins clean, but there are, as I mentioned in my paper, a large number of these units that are running for long periods of time between regenerations. There are going to be periods of business inactivity in the years to come when these units will run for weeks at a time without backwashings. During those periods it is important to keep bacteria and algae out of the resin beds. It is even more important with these beds of resins than it was with conventional zeolites in years gone by. We get fouling of the beds, try as we may to keep them clean with regenerants and an occasional treatment with chlorine.

Erickson: Will prior treatment with fungicides affect the performance of these exchangers in any way?

Alling: I don't think so for the reason that we are talking about almost infinitesimal traces of bacteria requiring only traces of pretreatment chemical. In most cases we have only a very small count of crenothrix, beggiatoa, asterionella, or other microscopic organisms in the water. In many cases the organisms have been destroyed by chlorination, but in such cases the cell structure or the protoplasm is left, and this coats over the resins and eventually reduces their capacity.

(The End)

Heat Transfer in Packed Beds

Prediction of Radial Rates in Gas-Solid Beds

W. B. Argo and J. M. Smith Purdue University, Lafayette, Indiana

An understanding of the fundamentals governing radial heat transfer in packed beds is important both when the bed is used as a heat-exchange device and when it serves as a catalytic reactor. In the design of packed-bed heat exchangers, usually it is not necessary to know point temperatures within the bed but only mean gas temperatures entering and leaving the system. In contrast, the temperature must be known as a function of radial position when the bed is employed as a nonadiabatic catalytic reactor. This is necessary in order to account for the effect of temperature upon the rate of reaction.

A method is proposed for predicting the effective thermal conductivity of a packed-bed heat exchanger or reactor by summing the contributions of each mechanism by which heat is transferred radially in the bed. Such thermal conductivities can be used to determine the temperature variation with radial position, which is of fundamental importance in the design of nonadiabatic catalytic reactors.

The equation developed includes the effects of void space, temperature level, mass velocity, particle size, thermal conductivity of the solid packing, and physical properties of the gas upon the effective thermal conductivity. Comparison with experimental data indicates good agreement between predicted and measured results where the conductivity applies to the bed itself. Over-all effective thermal conductivities include the effect of a wall resistance, and, hence, are lower than those applicable for computing temperature profiles within the bed.

Experimental Investigations

Because of the difference in requirements for the two applications just mentioned, investigations of radial heat-transfer rates have followed two different methods of approach. In the first, only boundary temperatures have been measured and the results correlated as heat-transfer coefficients h_w (based upon the difference between the bulk temperature of the gas flowing through the bed and the wall temperature) or over-all effective thermal conductivities k_{e_0}. The works of Colburn (6), Leva and co-workers (14-16), Hougen and Piret (10), Vershoor and Schuit (23), and Singer and Wilhelm (21) are of this type.

The second method is based upon the direct measurement of radial temperature traverses within the bed. Coberly and Marshall (5) and Felix and Neill (8) used high-velocity thermocouples placed across a diameter of the bed just above the top of the packing, while Bunnell et al. (4), Irvin, Olson and Smith (13) and Schuler, Stallings, and Smith (19) inserted thermocouple junctions into the solid pellets making up the packed bed. Results in each case were reported in terms of effective thermal conductivities (k_e) of the gas-solid bed. The knowledge of temperature traverses permitted evaluation of k_e as

a function of radial position and it was found that the resistance to heat transfer increased greatly near the wall. Coberly and Marshall, and Felix and Neill accounted for this by postulating that an additional resistance to heat transfer existed at the wall over that in the bed proper. These investigators could reproduce their measured temperature traverses by utilizing a wall heat-transfer coefficient and a constant effective thermal conductivity throughout the bed. Bunnell (4), Irvin (13) and Schuler (19) handled the problem by allowing the effective thermal conductivity to vary in a continuous manner as the wall was approached rather than postulate an additional resistance operating right at the wall.

Because of the wall resistance it is not possible to make an exact comparison between over-all effective thermal conductivities k_{e_0} and effective thermal conductivities applicable within the bed. Over-all values can be looked upon as average conductivities weighted in favor of the low results existing near the wall. These over-all conductivities may be used to predict the bulk mean temperatures of the gas leaving the bed but not the point temperatures necessary for reactor design.

Effect of Radial Velocity Variations

In addition to decreasing near the pipe wall, the effective thermal conductivities determined by Irwin and Schuler as a function of radial position showed a maximum value between the center and the wall. This result was also noted by Felix and Neill. Three recent investigations (1, 17, 20) of variations in mass velocity of gas with radial position in packed beds indicate a similar maximum flow rate within the bed, with a rapid drop as the pipe wall is approached. That this velocity variation is also the cause of the variation in effective thermal conductivity seems clear when the mechanisms by which heat is transferred radially are examined. Except for packing materials of high thermal conductivity, only a small fraction of the heat transferred radially is by conduction in the gas and solid phases in the bed. The major portion is the result of convection in the gas, or convection in the gas combined with conduction in the solid. Both mechanisms depend upon the degree of turbulence within the bed and hence upon the gas velocity. A thorough understanding of the transfer of heat and

mass radially will then depend upon a knowledge of the velocity pattern in the bed.

Mechanisms of Radial Heat Transfer

In the development of a method of predicting effective thermal conductivities two possibilities are available. One is to attempt an empirical or semi-empirical correlation in terms of all the properties of the gas and bed that are involved. Since heat is transferred by several mechanisms, this approach would not only be difficult but also would lead to results which could not be safely extrapolated beyond the range of existing data. A more fundamental approach is to correlate the contribution of each mechanism in terms of the relatively few properties that affect it and then combine these various contributions.

This second proposal was applied first by Schumann and Voss (18) and Damkoehler (7) and later modified by Wilhelm, Johnson, Wyncoop and Collier (22), particularly to static systems (no gas flow in the bed). Verschoor and Schuit (23) divided the total heat transfer into two parts by expressing the effective thermal conductivity as the sum of a static contribution and a contribution due to the flowing gas. Singer and Wilhelm (21) carried the development considerably further by postulating separate mechanisms by which heat is transferred radially in the bed. These mechanisms were: 1) a conductive process in the fluid phase, 2) a convective process in the fluid phase, 3) a conductive process from particle to particle through point-to-point contact and stagnant fluid in the zone of solid-to-solid contact, and 4) heat flow between fluid and solid particles. To these processes, Schuler later added radiation. In utilizing this approach experimental thermal data on the packed bed are necessary in order to evaluate the contribution of mechanism 3) for high conductivity particles.

It is apparent that Singer and Wilhelm considered the heat flow between fluid and solid particles to be intimately related to subsequent conduction in the solid phase, including conduction from one particle to the next through point contacts and stagnant gas fillets. No provision is made for heat entering and leaving the same particle by flow from and to the gas stream. Instead it is implied that after heat reaches a solid particle by mechanism, 2) further transfer must be through a solids path, by conduction through point contacts or stagnant gas fillets. This led Singer and Wilhelm to include point contact and gas filled resistances in their solid path, i.e., mechanism 3).

Leva and Grummer, in attempting to explain the improved heat-transfer char-

acteristics of beds packed with high-conductivity metals, suggested that some heat was transferred from the gas stream to an individual particle, transmitted through the particle by conduction, and subsequently removed from the same particle on the other side by transfer back to the gas stream. Analysis of this possibility indicates that this series-type of mechanism is an important contribution, even in the case of packing materials considered to be poor conductors. In other words, the path of lowest resistance for transfer of thermal energy already in a solid particle is not necessarily by conduction to the adjacent particle through point contact or stagnant gas fillets, but may be by conduction to the surface of the solid and out into the gas phase by an effective film resistance. The purpose of this paper is to present a quantitative explanation of radial heat transfer which takes both paths, as well as radiation, into account, and from which effective thermal conductivities may be predicted without experimental data on radial heat transfer with flowing gases.

Theoretical Development

At steady state conditions the heat transferred through a cylindrical plane parallel to the center line of a packed cylindrical bed will be the sum of the part passing through the void space and the part passing through the solid material. If q is this total rate of heat flow per unit area of the plane, the point effective thermal conductivity k_e is defined by the expression

$$q = -k_e \frac{\partial t}{\partial r} = q_{void} + q_{solid} \qquad (1)$$

This concept is illustrated in the upper half of Figure 1. The temperature gradient in Equation (1) applies to the bed as a whole and deserves some explanation because of the possibility of temperature differences between solid and fluid phases in a packed bed. Bunnell measured temperatures at the same radial position in both the gas and in the center of the solid particle and found no significant difference. The solid particles were activated alumina, a material of relatively low conductivity. On the basis of these data it will be assumed that the average temperature of the particle is the same as that of the gas at the same radial position. This does not require that the temperature gradient within a single solid particle coincide with that of the fluid phase.

As an illustration, Bunnell measured effective thermal conductivities of the order of 0.10 to 0.30 B.t.u./(hr.)(sq.ft.) (° F.) while the conductivity of his alumina pellets is estimated to be about 0.5. Clearly the solid particles are not conducting heat under the gross radial

gradient observed for the bed as a whole. For solid packings with thermal conductivities of 0.5 or greater, the gradients within the pellets would normally be less than those for the whole bed. This situation is shown graphically in the lower portion of Figure 1. The higher the true thermal conductivity of the solid the less the temperature change in the solid phase, so that with metallic pellets the temperature within a single particle would be expected to approach a constant value. The gradient in Equation (1) is the value for the bed as a whole and is designated as the observed gradient in Figure 1. It is the same as that in the continuous fluid phase.

As shown in Figure 1, there may exist a considerable difference in temperature between gas and particle for heat transfer to the particle on one side and for heat transfer from the particle back to the gas stream on the other side. The temperature gradient within the particle would then be just sufficient to transfer this heat from one side to the other.

The heat passing across the plane in the void space is the sum of that due to molecular conduction, turbulent diffusion, and radiation. Since these paths are in parallel,

$$q_{void} = -\delta(k'_c + k'_{td} + k'_r) \frac{\partial t}{\partial R} \qquad (2)$$

where δ is the void fraction and the prime superscript indicates that the conductivity is based upon the total void and nonvoid area: i.e., $k_r = \delta k'_r$, etc.

The precise evaluation of the heat transfer through the particle presents a mathematically complex problem. This complexity arises both from the geometry of the bed and the several mechanisms by which heat can enter and leave the pellet. To solve this problem, no simplification will be made concerning the heat-transfer mechanisms, but an ideal model of the packed bed will be employed in order to avoid unsolvable geometrical difficulties. The methods by which heat can enter a particle from its inner side are radiation, convection from the gas stream, and conduction through point contacts and stagnant fillets as indicated in Figure 2. Heat is transferred through the particle and leaves the other side by the same three mechanisms. The three processes are in series and hence the whole will be designated as the series mechanism. Hence

$$q_{solid} = -k'_{series}(1-\delta) \frac{\partial t}{\partial R} \qquad (3)$$

Combining Equations (1), (2), and (3) gives an expression for the point effective thermal conductivity in terms of contributions for each mechanism responsible for radial heat transfer.
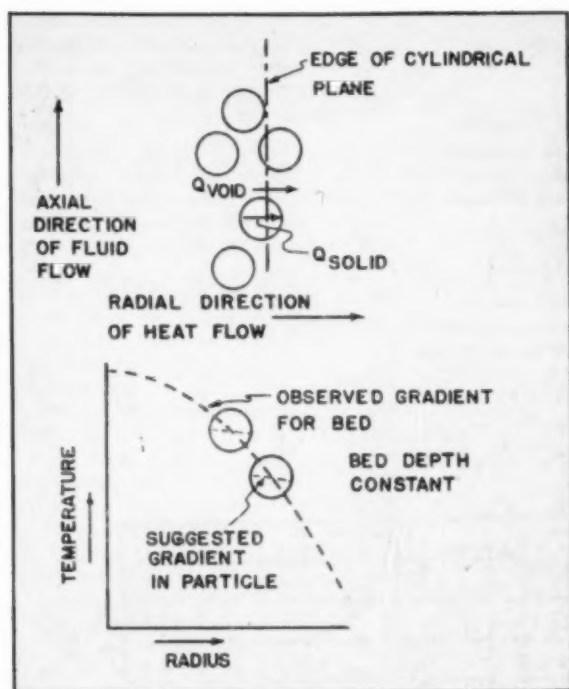
Fig. 1. Heat flow through packed beds.
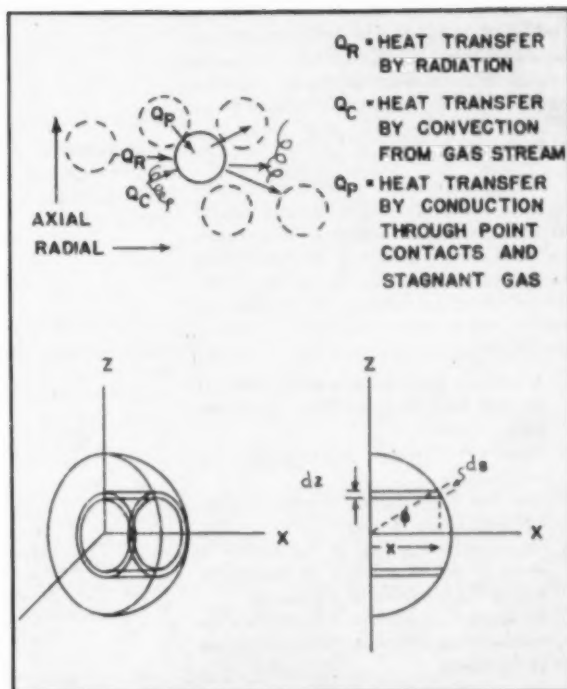
Fig. 2. Heat transfer in a single spherical particle.

$$k_e = \delta(k'_e + k'_{td} + k'_r) + (1 - \delta)k'_{series} \qquad (4)$$
$$= k_e + k_{td} + k_r + k_{series}$$

FLUID PHASE CONDUCTION

The value of k'_e in Equation (4) is the molecular conductivity of the fluid. Its value will change with radial position in the bed because of temperature variation. For gases k_e is so low that it is not an important contribution to k_e, while for liquids this is not true.

TURBULENT DIFFUSION

The contribution of turbulent diffusion k_{td} is a measure of heat transfer as a result of turbulent mixing of portions of the gas stream at different temperatures. As Singer and Wilhelm have pointed out, its value can be advantageously estimated from measurements of mass transfer radially by the same mechanism. The advantage of using mass-transfer data is due to the fact that the transfer of mass radially in a packed bed does not involve the series or radiation mechanisms, but is caused only by molecular conduction and turbulent diffusion and the former contribution is small. On this basis,

$$k'_{td} = \rho c_p D_e / \delta$$

or in terms of modified Peclet numbers

$$\left(Pc_m = \frac{d_p u}{D_e}\right)$$

determined from mass-transfer data,

$$k'_{td} = \rho c_p \left(\frac{d_p u}{\delta Pc_m}\right) = \frac{d_p c_p G}{Pc_m \delta} \qquad (5)$$

Bernard and Wilhelm (3) have presented data for mass-transfer Peclet numbers for one gas system—air flowing through packed beds of ⅜-in. spheres in an 8-in. pipe. This information is presented in Figure 3 and may be used as an approximate means of obtaining the turbulent diffusion contribution. At high velocities Baron (2) has shown on theoretical grounds that the Peclet number for mass transfer should be independent of particle size. However, additional mass-transfer data using different pellet sizes and shapes are necessary before k_{td} can be computed with full confidence for any packed bed conditions. In the comparisons of predicted and experimental k_e values given in the next section, Bernard and Wilhelm's results are used for all packings because of the absence of additional data.

The contribution of turbulent diffusion may vary significantly with radial position especially near the pipe wall where the void fraction and velocity are varying. Results in Figure 3 are average values determined for the bed as a whole, but, of course, do not include the wall resistance that is associated with over-all effective thermal conductivities.

RADIATION CONTRIBUTION

The value of k'_r depends upon the temperature level and gradient. Hence, it will vary with radial position if the temperature gradient is large. An accurate evaluation of the radiation contribution requires experimental temperature measurements. Schuler, Stallings and Smith, using measured temperature gradients, found (1) that k'_r reached a maximum at 0.4 to 0.6 of the distance from the center to the wall and (2) that the temperature level was more important in establishing k'_r than the temperature gradient. Thus, Damkoehler's simplified expression for radiation (corrected for emissivities < 1.0)

$$k'_r = 4\left(\frac{\epsilon}{2 - \epsilon}\right) d_p (0.173) \left(\frac{T_a^3}{100^4}\right) \qquad (6)$$

was found to agree well with more elaborate methods, even though it included only the bulk mean temperature of the bed, T_a (in ° R. in Eq. (6)).

Schuler's investigations were made at temperatures as high as 400° C. and with large temperature gradients. At low mass velocities the radiation contribu-

tion to k_e was of the order of 10-15%. Hence, an approximate equation such as Equation (6) is satisfactory for estimating k'_r for most practical cases of heat transfer or reaction in packed beds.

Series Mechanism

The mathematical problems encountered in evaluating the series mechanism in an actual packed bed with its non-uniform arrangement of solid particles have been mentioned. In order to obtain a useful solution to the problem of radial heat transfer through the pellet, the following assumptions will be made:

1. A constant temperature gradient exists in the fluid phase for a radial distance of one pellet diameter.
2. A constant temperature gradient exists in the solid particle.
3. Heat flow in the particle is radial in direction, with respect to reactor.
4. The mean temperature of the particle is equal to the temperature of surrounding fluid at the center line of the particle.
5. An average heat-transfer coefficient may be considered applicable to the entire surface of the particle.

Assumption (3) permits the heat transfer in the particle to be summed over elements normal to the direction of flow. One half of such a spherical particle with the normal element is illustrated in the lower half of Figure 2. An energy balance on the element leads to the expression

$$dq' = k_s(2\pi z dz)\left(\frac{t - t_s}{x}\right)$$
$$= h 2\pi z (t_s - t_g) ds \qquad (7)$$

dq' = rate of heat flow through the element

k_s = molecular thermal conductivity of solid particle

z = coordinate normal to heat flow (parallel to axis of bed)

x = coordinate in direction of heat flow

t = temperature at the center plane of the particle = temperature of fluid phase at the same radial position

t_s = temperature of surface of the particle at a distance x

$t_g = t + x\dfrac{dt}{x}$ = temperature of the fluid at a distance x

ds = element of surface through which heat is transferred to the gas stream or to other particles

h = total heat-transfer coefficient from the surface of the particle to the fluid or to other particles

The element of surface area available for heat transfer to the gas stream may be related to dz as follows:

$$ds = \frac{dz}{\cos \phi} = \frac{r dz}{x} \qquad (8)$$

where r is the radius of the sphere.

Elimination of t_s from Equation (7) and integration to obtain the total heat flow through the central plane of the particle yields:

$$q' = -\frac{hk_s\pi}{(k_s + hr)}(r^3)\left(\frac{dt}{dx}\right) \qquad (9)$$

At any distance x from the central plane the total heat flow in the particle is

$$q' = -\frac{hk_s r}{k_s + hr}(\pi z^2)\left(\frac{dt}{dx}\right) \qquad (10)$$

Since πz^2 is the cross-sectional area of the solid at this point, and since dt/dx is the radial temperature gradient in the fluid phase, Equation (10) may be written,

$$q' = -\left(\frac{hk_s r}{k_s + hr}\right)A\,\frac{dt}{dR}$$
= heat transferred through the solid, based upon the temperature gradient of the bed (11)

Since q_{solid} is based upon unit area of normal plane, including both void and nonvoid surface,

$$q_{solid} = \frac{q'}{A}(1-\delta) = -\left(\frac{hk_s r}{k_s + hr}\right)$$
$$(1-\delta)\,\frac{\partial t}{\partial R} \qquad (12)$$
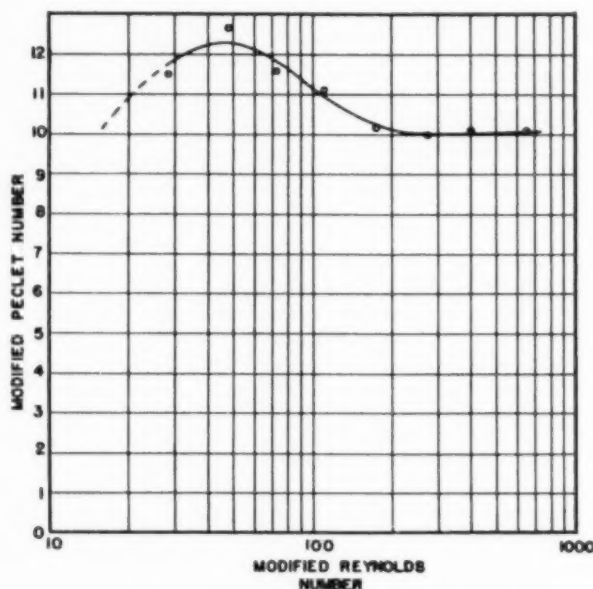
Fig. 3. Mass transfer data of Bernard and Wilhelm (3).

Fig. 4. Effective thermal conductivities, ⅛-in. cylinders.
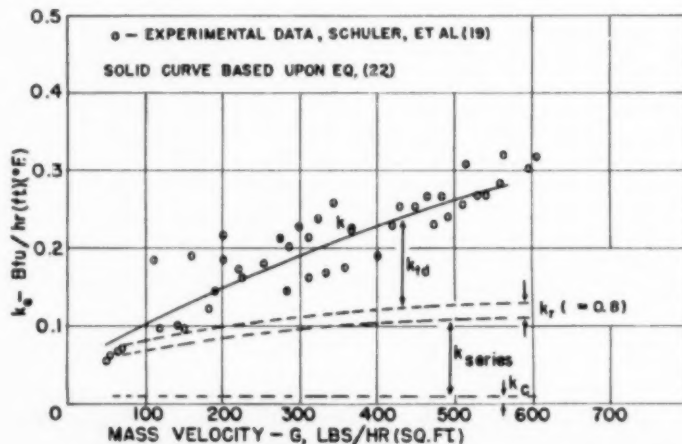
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Comparison of Equation (3) and (12) gives the desired expression for k'_{series}.

$$k'_{series} = \frac{hk_s r}{k_s + hr} = \frac{hk_s d_p}{2k_s + hd_p} \quad (13)$$

Similar derivations for cylindrical particles oriented in the bed with the cylindrical axis either perpendicular or normal to the axis of the pipe yield exactly the same result as Equation (13), where d_p is the length of the cylinder if it is oriented normal to the axis of the pipe. In order to generalize Equation (13) for random orientations of cylinders and for other shaped particles, it is suggested that d_p be taken as the diameter of the sphere with the same surface area as the particle.

The evaluation of k'_{series} requires a knowledge of the total heat-transfer coefficient h. This will be the sum of the convection coefficient from the fluid surrounding the particle, the radiation contribution from adjacent particles, and a conduction contribution from pellets in contact, with the particle. Thus h may be defined as follows:

$$h = \frac{q'}{A(\Delta t)_m} = \frac{q'_c + q'_r + q'_p}{A(\Delta t)_m}$$
$$= h_c + h_r + h_p \quad (14)$$

The convection coefficient h_c can be predicted from the data on heat transfer between solids and fluids in packed beds, such as the work of Hougen and Wilke (12) and Hougen, Gamson, and Thodos (11). Their correlations are:

$$\left(\frac{h_c}{c_p G}\right)\left(\frac{c_p \mu}{k}\right)^{\frac{2}{3}} = 1.95 \left(\frac{d_p G}{\mu}\right)^{-0.51}$$
$$Re < 350 \quad (15)$$

$$\left(\frac{h_c}{c_p G}\right)\left(\frac{c_p \mu}{k}\right)^{\frac{2}{3}} = 1.06 \left(\frac{d_p G}{\mu}\right)^{-0.41}$$
$$Re > 350 \quad (16)$$

The radiation and conduction coefficients, h_r and h_c depend upon the value of $(\Delta t)_m$ defined by Equation (14). A derivation based upon the same assumptions as employed in obtaining Equation (13) leads to the result

$$(\Delta t)_m = -\frac{d_p k_s}{2(2k_s + hd_p)} \frac{dt}{dR} \quad (17)$$

Since Equation (6) gives the radiation contribution in terms of k'_r, h_r must be expressed in terms of k'_r

$$h_r = -k'_r \left(\frac{A'}{A}\right) \frac{\frac{dt}{dR}}{(\Delta t)_m} \quad (18)$$

where

A' = projected area of one half of spherical particle = $\pi d_p^2/4$
A = area of one half of spherical particle = $\pi d_p^2/2$

Combination of Equation (17) and (18) to eliminate $(\Delta t)_m$ gives for h_r

$$h_r = \frac{k'_r(2k_s + hd_p)}{d_p k_s} \quad (19)$$

Equation (19) permits the evaluation of h_r by employing Equation (6) for k'_r.

The analogous expression for h_p is

$$h_p = \frac{k'_p(2k_s + hd_p)}{d_p k_s} \quad (20)$$

The conductivity k'_p determines the heat transfer through the pellet by point contact and through stagnant gas fillets surrounding the contact point. The correlation of Wilhelm, Wynkoop, Johnson, and Collier, summarized by Equation (21), may be used with Equation (20) to determine h_p.

$$\log_{10} k'_p = 1.76 + 0.0129 k_s/\delta \quad (21)$$

Summation of the expressions for each mechanism in accordance with Equation (4) results in a composite equation for the effective thermal conductivity

$$k_e = \delta \left[k'_c + \frac{d_p c_p G}{Pe_m \delta} + 4\left(\frac{\epsilon}{2-\epsilon}\right) d_p (0.173)\left(\frac{T_a^3}{100^4}\right)\right] + (1-\delta)\frac{hk_s d_p}{2k_s + hd_p} \quad (22)$$

in which the heat-transfer coefficient h is given by Equations (14), (15), (19), and (20). Equation (22) predicts the effect of a number of the basic variables upon k_e. Since Pe_m is rather insensitive to particle diameter, the second, third and fourth terms in the equation require that k_e increase with particle diameter. Since h increases with mass velocity G, the Equation predicts that k_e will increase with G. It is expected also that k_e will increase with the conductivity of the solid particle although the increase will be slight because k_s occurs in both numerator and denominator of the last term. All these effects are verified by the available experimental data, which are examined in the next section.

When the temperature in the bed is less than 300° C., which is frequently the case, the radiation contributions to k_e may be neglected. Under these circumstances, Equation (22) can be simplified to the following form:

$$k_e = \delta(k'_c + k'_{td})$$
$$+ (1-\delta)k_s \frac{d_p h_c + 2k'_p}{d_p h_c + 2k_s} \quad (23)$$

where k'_p is the conductivity between solid particles in contact and is given by Equation (21). If the Peclet number for heat transfer is introduced through the relation $Pe_h = d_p c_p G/k_e$, and Equation (15) is used for h_c, Equation (23) may be written,

$$\frac{1}{Pe_h} = \delta \left[\frac{1}{(Pr)(Re)} + \frac{1}{\delta Pe_m} \right]$$

$$+ (1-\delta)\frac{k_s}{(Re)\mu c_p}$$
$$\left[\frac{1 + \frac{k'_p(Re)^{-0.49}(Pr)^{\frac{2}{3}}}{c_p \mu}}{1 + \frac{k_s(Re)^{-0.49}(Pr)^{\frac{2}{3}}}{c_p \mu}} \right] \quad (24)$$

This expression shows that if k_e is correlated in terms of a Peclet number for heat transfer, that a plot of Pe_h vs. Re will be independent of particle diameter and mass velocity, but not of void fraction and particle conductivity, k_s (note that the contact conductivity, k'_p, is, according to Equation (21), a function of δ and k_s and therefore is not an independent variable in Equation (24)). Figure 10 shows a plot of Pe_h vs. Re predicted from Equation (24) for $\delta = 0.40$ and two values of k_s. The upper curve is based upon $k_s = 0.5$ which is a typical value for ceramic materials and the lower for $k_s = 19$, which is an example of a metallic packing, i.e., lead. The points on the figure refer to Pe_h values based upon over-all, effective thermal conductivities determined experimentally by Singer and Wilhelm. These will be discussed in the next section.

Application of Equations

To illustrate the application of Equation (22) to the calculation of effective thermal conductivities and to show the relative magnitude of the several contributions, values of k_e are predicted for two specific cases for which experimental results are available for comparison. For each the following data apply:

Gas — Dry air.
Pipe — 2-in. I.D., with a constant wall temperature of 200° C.
Particle — ⅛-in. cylindrical, alumina pellets; $k_s = 0.5$ B.t.u./(hr.)(° F.)(ft.)
Void fraction — 0.40

Calculations are summarized in Table 1. Neglecting radiation contributions causes a difference of 17% in Case I but only 6% in Case II where the mass velocity is higher. It should be noted that the contribution of turbulent diffusion is but 28% in Case I, while it has increased to 52% for Case II.

This procedure for predicting k_e values can be used for any conditions of packing and gas flow, provided the correlation for each individual mechanism is valid under those conditions. Probably the area of greatest weakness is in the correlation for the contribution of turbulent diffusion. It will be

Table 1.—Estimated Effective Thermal Conductivities (Eq. (4))

B.t.u./(hr.)(ft.)(° F.)

	Case I $G_o = 147$ lb./ (hr.)/sq.ft.)	Case II $G_o = 493$
Mean bed temperature—° F.	455	536
k'_r, radiation conductivity from Eq. (6)	0.029	0.029
k'_p, point contact conductivity from Eq. (21)	0.018	0.018
h_r (Eq. (19))	5.30	3.50
h_p (Eq. (20))	3.33	3.33
h_o (Eq. (15))	15.6	28.3
h, sum of h_r, h_p and h_o	24.2	36.9
k'_{series} (Eq. (13))	0.115	0.168
$k_r = \delta k'_r$	0.010	0.010
$k_r = \delta k'_r$	0.012	0.012
k_{td}, turbulent diffusion contribution (Eq. (5))	0.035	0.135
k_e (Eq. (4))	0.126	0.258
k_e (neglecting radiation)	0.104	0.241
k_e, experimental (Schuler et al., (19))	0.113	0.276
Modified Reynolds number	25.4	85.2

recalled that the use of Equation (5) is restricted to the data of Bernard and Wilhelm for ⅜-in. cylindrical packing through which air was flowing. When information for other systems of gases and pellet sizes are available more appropriate values of the Peclet number can be used in Equation (5) to predict k_{td}.

Experimental results shown in Table 1 are the average of k_e values reported by Schuler et al., for all radial positions. Similarly, the calculated thermal conductivities are based upon the average mass velocity and void fraction for the entire bed. The proposed equations for determining k_e are satisfactory for computing point values of effective thermal conductivities. However, the experimental data used to evaluate k_{td} (3) and h_e (12) in the equations were determined for the average conditions of the bed. Until point values of k_{td}, h_e, and void fraction are available for packed beds, the prediction of point values of effective thermal conductivities is hazardous.

The average values of k_e referred to in Table 1 should not be confused with the over-all values determined from heat-transfer data such as reported by Leva, Wilhelm and Singer, and Colburn. As mentioned earlier these over-all effective thermal conductivities include a wall resistance.

Comparison with Experimental Data

a. Schuler, et al. (19).

On the basis of experimental temperature measurements Schuler determined effective thermal conductivities for ⅛-in., ³/₁₆-in. and ¼-in. cylindrical pellets packed in a 2-in. I.D. pipe through which air was flowing. These k_e values were computed by taking into account radial variations in both mass velocity and k_e itself,

and, as such, are point conductivities. Figures 4, 5, and 6 show a comparison of the experimental data and the predicted values (solid curves) based upon Equation (4). As Schuler pointed out these charts show that the variation in k_e with radial position is taken into account, at least in part, by the radial changes in mass velocity. The scatter of the points is due to experimental limitations and also to the residual influence of radial position, not correlated by the variation in mass velocity. This residual effect is probably explainable in terms of variations in void fraction with radial position, particularly near the wall of the pipe.

The agreement between the predicted curve and the data is considered to be good in view of the fact that Equation (4) requires for its application no radial heat-transfer data. It is

to be noted that the theory predicts the effect of particle size on the effective thermal conductivity. Figure 7 combines the results of Figures 4, 5, 6 in the form of Peclet numbers $(d_p Gc_p/k_e)$ for heat transfer plotted vs. modified Reynolds number. The experimental points shown are average values for all radial positions.

b. Felix and Neill (8), Coberly and Marshall (5).

The only other experimental data for effective thermal conductivities that do not include a wall resistance are those of Coberly and Marshall and Felix and Neill. These investigations were made in the same apparatus and agree well with each other. However, the results of both yield k_e values which average about 100% higher than those obtained by Schuler and Bunnell, and predicted by the theoretical methods described in this paper. Table 2 gives a comparison of some of Coberly and Marshall's results and those computed from Equation (23). The average temperature (160° F.) was sufficiently low that radiation could be neglected so that a contribution for k'_r, was not employed.

It is possible to analyze this discrepancy by comparison with the mass-transfer data of Bernard and Wilhelm in the following way: The experimental k_e values for each investigation can be reduced by the contributions due to radiation, molecular conduction, and the series mechanism in accordance with Equation (4). This yields a value of k_{td}, and hence Pe_m, which may be compared with Bernard and Wilhelm's experimental values. Results are shown in Figure 8 for particle sizes as nearly the same as the data permit. The subtraction of two numbers to obtain a smaller k_{td} is not a precise procedure and the points shown on the figure scatter. However, it is clear that the results of Coberly and Marshall are not in agreement with those of Bernard and Wilhelm. As a result of omission of the void fraction, Felix and Neill predicted

Table 2.—Comparison of Predicted k_e Values (Eq. (22)) and Experimental Results of Coberly and Marshall (5)

G lb./(hr.)(sq.ft.)	Re	$(1 - \delta)k'_{series}$	k_{td}	k_r	k_e predicted	k_e Exp—(5)
⅛ in. × ⅛-in. Celite cylinders						
175	44.6	0.041	0.044	0.007	0.091	0.30
310	79.0	0.046	0.081	0.007	0.134	0.42
475	121	0.050	0.135	0.007	0.185	0.50
750	191	0.054	0.225	0.007	0.286	0.60
¼ in. × ¼-in. Celite cylinders						
175	89.2	0.047	0.093	0.007	0.147	0.32
310	158	0.052	0.183	0.007	0.242	0.38
475	242	0.056	0.291	0.007	0.354	0.57
750	382	0.060	0.460	0.007	0.527	0.90
1005	512	0.063	0.617	0.007	0.686	1.28
1215	619	0.064	0.745	0.007	0.817	1.27
⅜ in. × ½-in. Celite cylinders						
475	401	0.060	0.483	0.007	0.550	0.97
750	633	0.065	0.762	0.007	0.827	1.31
1005	848	0.067	1.022	0.007	1.096	1.88
1215	1026	0.069	1.235	0.007	1.315	2.01

Average temperature = 160° F.
$k_e = 0.143$ B.t.u./(hr.)(ft.)(° F.)

Radiation neglected
$\delta = 0.40$
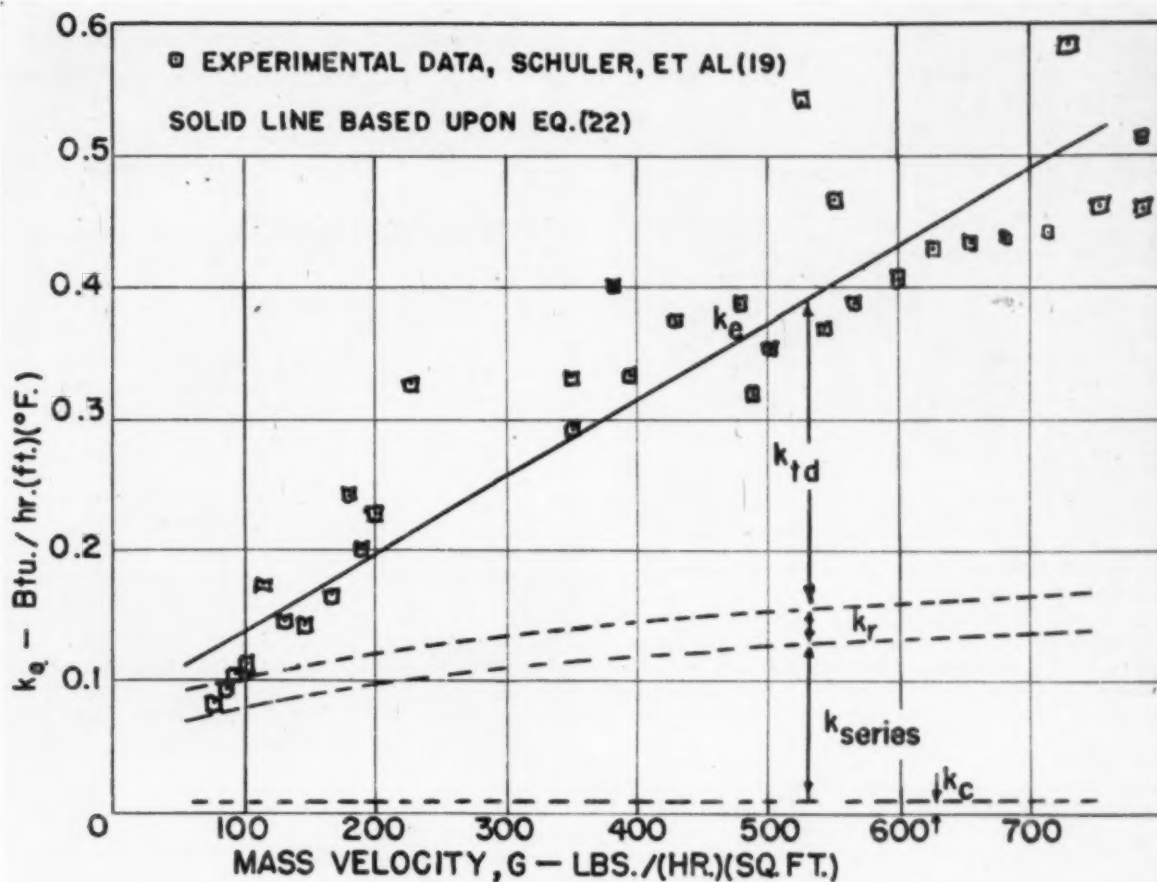
Fig. 5. Effective thermal conductivities, 3/16-in. cylinders.
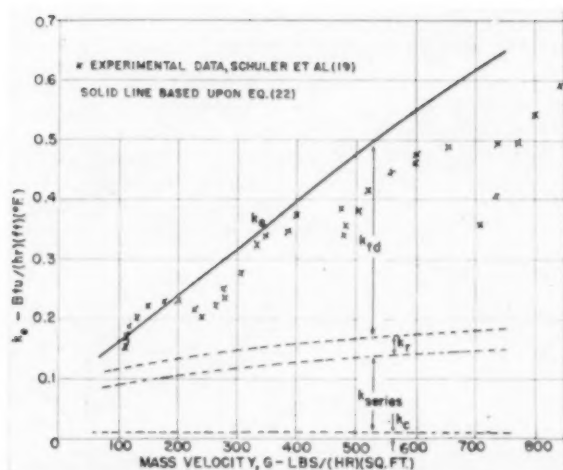
Fig. 6. Effective thermal conductivities, ¼-in. cylinders.
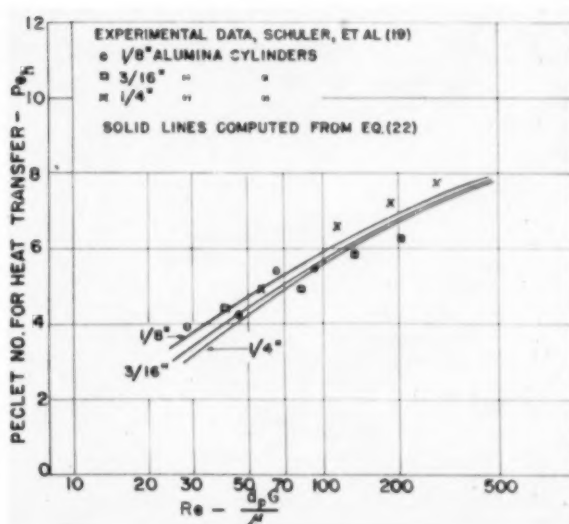
Fig. 7. Pe vs. Re for alumina cylinders (radiation included).
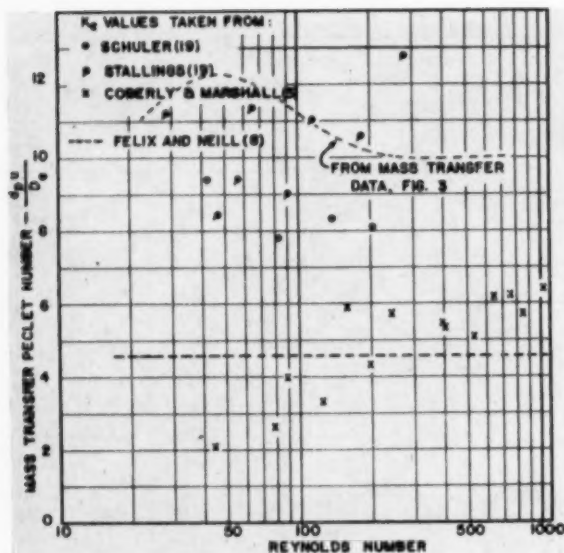
Fig. 8. Peclet numbers based on experimental k_e values.

Fig. 9. Effect of solid conductivity on k_e. ▶
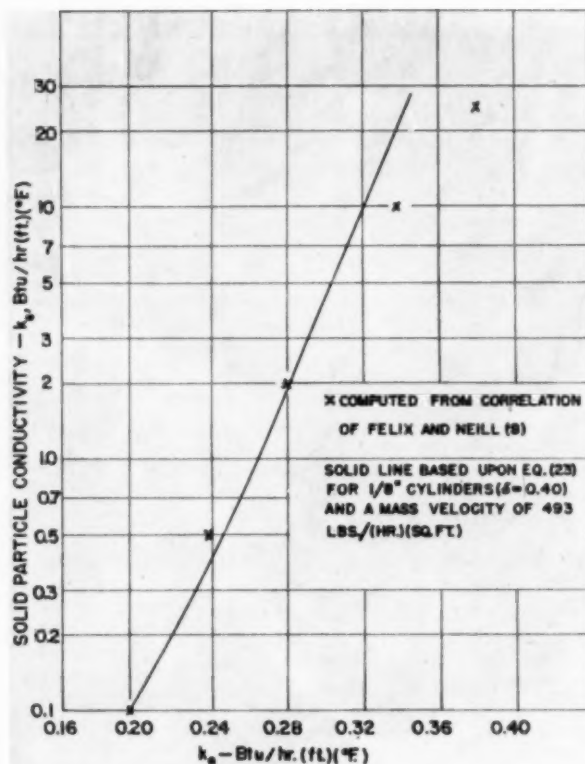
Fig. 10. Comparison of overall and bed effective thermal conductivities.
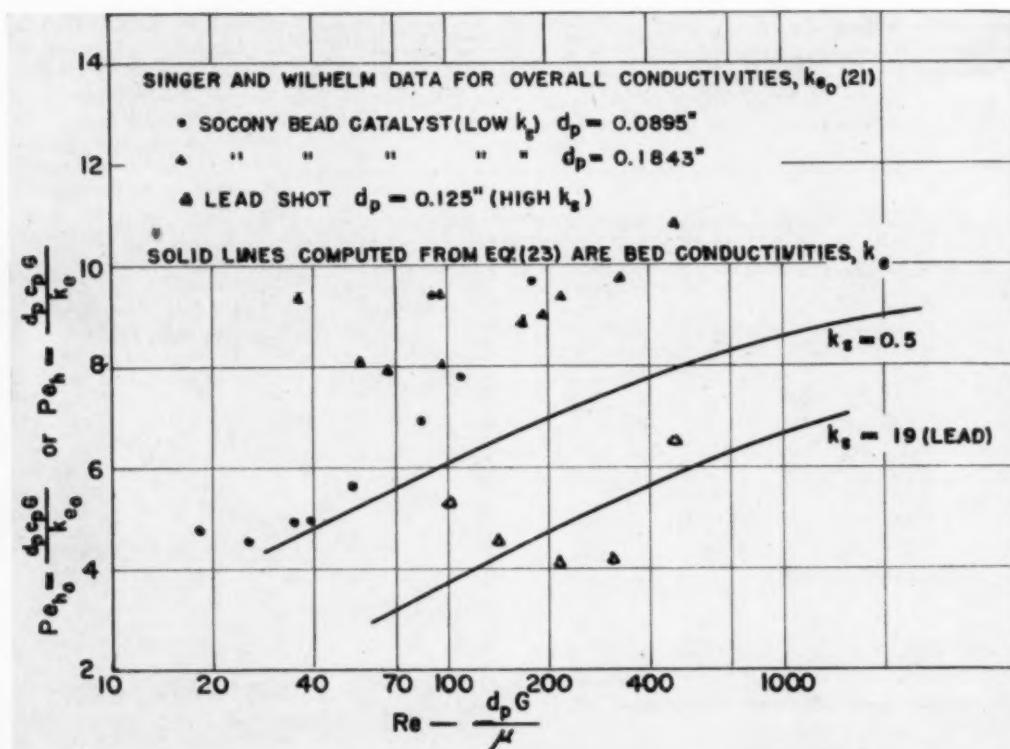
agreement of their results with those of Bernard and Wilhelm. However, if the equation for their data is corrected using a void space of 0.40, it becomes

$$P_{e_m} = \frac{d_p u}{D_e} = 4.6$$

which is considerably less than Bernard and Wilhelm's data, as indicated in Figure 8. It is possible that the high velocity thermocouple assembly used in the apparatus may have caused excessive mixing and turbulence, thus yielding low values of the Peclet number.

c. Effect of Pellet Conductivity.

Felix and Neill investigated the effect of thermal conductivity of the solid pellet k_s and correlated the data by the following equation:

$$k_{e_2} = k_{e_1}\left(\frac{k_{s_2}}{k_{s_1}}\right)^{0.32} \tag{25}$$

Although there is some doubt concerning the absolute magnitude of their effective thermal conductivities, it should be possible to use Equation (25) to compare the approximate effect of k_s on k_e with that predicted by the equations proposed in this paper. Figure 9 shows k_e values plotted vs. k_s for ⅛-in. cylindrical pellets through which air is flowing at a mass velocity of 493 lb./(hr.)(sq.ft.). The solid line is that predicted from Equation (23) (using a void fraction of 0.40), while the points represent values computed from the Felix and Neill correlation. In using Equation (25), k_e was chosen equal to the predicted value at $k_s = 0.1$ B.t.u./(hr.)(ft.)(° F.).

d. Comparison with Over-all Effective Thermal Conductivities.

While the over-all effective thermal conductivity k_{e_0} is not equivalent to the k_e predicted by Equation (23), it is interesting to compare the two types of conductivities. Values of the Peclet number ($P_{e_h} = d_p c_p G / k_{e_0}$) are shown as points on Figure 10 for Socony bead catalyst and for lead spheres determined experimentally by Singer and Wilhelm. The over-all Peclet numbers for the low-conductivity Socony bead catalyst are in general higher than the predicted conductivities for the bed. This is expected since the wall resistance reduces k_{e_0}. For the high-conductivity particles, $k_s = 19$ B.t.u./(hr.)(ft.)(° F.), the over-all values are about the same as the predicted results. Since the wall resistance primarily affects $k_{t d}$ and since k_{series} is relatively large for high-conductivity solids, k_{e_0} and k_e should be more nearly identical. It will be noted that the predicted k_e curves and the experimental k_{e_0} results are for approximately the same size particles. For large values of d_p, the Peclet numbers based upon k_{e_0} would increase because of the increased importance of the wall resistance.

Conclusions

A method has been proposed for predicting effective thermal conductivities in packed beds by summing the contribution of each mechanism by which heat is transferred radially in the bed. These mechanisms are: 1) conduction in the gas phase, 2) convection in the gas phase, 3) radiation, and 4) a series mechanism accounting for the heat transfer from solid particle to particle by conduction, convection, and radiation. Comparison with the data available indicates that the theory satisfactorily predicts the effects of mass velocity, particle diameter, and conductivity of the solid particle on radial heat transfer in packed beds.

Notation

A = heat-transfer area in radial direction

c_p = specific heat at constant pressure

d_p = equivalent pellet diameter

D_e = effective diffusivity of gas in packed bed, based upon void plus nonvoid area

G = superficial mass velocity of gas through packed bed

h = over-all heat-transfer coefficient between gas and solid particle in packed bed

h_c = convection heat-transfer coefficient between gas and particle

h_p = equivalent heat-transfer coefficient for contact conductivity between solid particles

h_r = equivalent heat-transfer coefficient for radiation to solid particle

k_e = effective thermal conductivity in bed

k_{e_0} = over-all effective thermal conductivity for heat transfer from bed to wall of pipe

k'_e = true thermal conductivity of gas based upon void area. If k_e is based upon void plus nonvoid area, $k_e = k'_e \delta$

k'_p = point-to-point contact conductivity of particles in bed

k'_r = equivalent radiation conductivity

k_s = true thermal conductivity of solid particle material

k'_{series} = conductivity of series mechanism based upon nonvoid area.
$k_{series} = (1 - \delta)k'_{series}$ where k_{series} is based upon void and nonvoid area

k'_{td} = conductivity in gas due to turbulent diffusion: $k_{td} = \delta k'_{td}$

Pe_h = Peclet number for heat transfer in packed bed $Pe_h = \dfrac{d_p G c_p}{k_e}$

Over-all $Pe_{h_0} = \dfrac{d_p G c_p}{k_{e_0}}$

Pe_m = Peclet number for mass transfer $Pe_m = \dfrac{d_p u}{\delta D_e} = \dfrac{d_p G c_p}{\delta k'_{td}}$

Pr = Prandtl number for gas, $c_p \mu / k'_e$

q = heat-transfer rate per unit area, q_{void} and q_{solid} refer to heat transfer through void and nonvoid areas

Re = Reynolds number, $d_p G / \mu$

R = radial distance from center of pipe

r = radius of particle (used as a subscript r denotes radiation heat transfer)

T_a = mean temperature of bed, ° R.

t = temperature, ° F.

u = superficial velocity of gas

Greek Letters

ϵ = emissivity of solid particles

δ = void fraction

μ = viscosity of gas

ρ = density of gas

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PLASTICS EQUIPMENT REFERENCE SHEET

RAYMOND B. SEYMOUR and ROBERT H. STEINER

The Atlas Mineral Products Co., Mertztown, Pa.

CHLORINATED RUBBER COATINGS: Protective coatings and primers based on solutions of chlorinated rubber together with appropriate plasticizers, pigments, stabilizers and other compounding ingredients.

APPLICATION AND REMARKS: Chlorinated rubber serves as a base for a wide variety of commercial protective coatings. Any conventional method of application, such as brushing, spraying or roll coating may be used on properly prepared metallic or masonry surfaces. Excellent adhesion can be obtained without the use of primers but specially formulated primers containing inhibiting pigments such as red lead or zinc chromate must be used where the coating is to be exposed to highly corrosive atmospheres such as acid fumes or salt spray. Good outdoor weathering resistances are obtained through the use of properly stabilized chlorinated rubber coatings. A minimum of three coats is recommended for protection against fumes, vapors, and occasional splash and spillage of corrosive materials. Thin organic coatings are not recommended for continuous immersion in corrosive media but may be used to prevent contamination of noncorrosive liquids.

CHEMICAL COMPOSITION: Chlorinated rubber is manufactured by reacting chlorine with natural rubber in organic solutions to form a product containing about 67% chlorine which has lost all its original elasticity. Because of the wide variation possible in the compounding of chlorinated rubber coatings, it is important that only coatings, formulated specifically for corrosion resistance, be used.

TEMPERATURE LIMITATIONS: Because of the inherent instability of chlorinated rubber at higher temperatures, coatings based on this material are not recommended for continuous service above 125° F.

A TYPICAL CHLORINATED RUBBER COATING

MECHANICAL AND PHYSICAL PROPERTIES

Specific gravity	1.65
Water absorption %	0.3
Flexibility	Good
Adhesion	Excellent
Impact resistance	Excellent
Abrasion resistance	Good
Weather resistance	Good

CORROSION RESISTANCE

ACIDS	C	H
Acetic, 10%	F	N
Acetic, glacial	N	N
Benzene sulfonic	E	E
Benzoic	E	E
Boric	E	E
Butyric	N	N
Chloroacetic	N	N
Chromic, 10%	G	N
Chromic, 50%	P	N
Citric	E	E
Fatty acids (C₆ and up)	N	N
Fluosilicic	E	E
Formic	N	N
Hydrobromic	E	E
Hydrochloric	E	E
Hydrocyanic	E	E
Hydrofluoric	E	G
Hypochlorous	G	E
Lactic	E	F
Maleic	G	F
Nitric, 5%	G	F
Nitric, 20%	F	N
Nitric, 40%	N	N
Oleic	N	N
Oxalic	E	F
Perchloric	E	N
Phosphoric	F	N
Picric	P	N
Stearic	F	F
Sulfuric, 50%	F	F
Sulfuric, 70%	G	F
Sulfuric, 93%	N	N
Oleum	N	N
Mixed acids, 28% HNO₃ 55% H₂SO₄	N	N

ALKALIES		
Ammonium hydroxide	E	E
Calcium hydroxide	E	E
Potassium hydroxide	E	E
Sodium hydroxide	E	E

ACID SALTS	C	H
Alum	E	E
Ammonium Cl, NO₃, SO₄	E	E
Copper Cl, SO₄	E	E
Ferric Cl, SO₄	E	E
Nickel Cl, SO₄	E	E
Stannic, Cl	E	E
Zinc Cl, SO₄	E	E

ALKALINE SALTS		
Barium sulfide	E	E
Sodium bicarbonate	E	E
Sodium carbonate	E	E
Sodium sulfide	E	E
Trisodium phosphate	E	

NEUTRAL SALTS		
Calcium chloride	E	E
Calcium sulfate	E	E
Magnesium Cl, sulfate	E	E
Potassium Cl, NO₃, SO₄	E	E
Sodium Cl, NO₃, SO₄	E	E

GASES		
Chlorine wet	F	N
Chlorine dry	N	N
Sulfur dioxide wet	E	E
Sulfur dioxide dry	E	E
Hydrogen sulfide	E	

ORGANIC MATERIALS		
Acetone	N	N
Alcohols, methyl, ethyl	E	E
Aniline	N	N
Benzene	N	N
CCl₄	N	N
Chloroform	N	N
Ethyl acetate	N	N
Ethylene chloride	N	N
Formaldehyde, 37%	E	N
Gasoline	P	N
Phenol	N	N

	C	H
Refinery crudes	N	N
Trichloroethylene	N	N

PAPER-MILL APPLICATIONS		
Kraft liquor	E	E
Black liquor	E	E
Green liquor	E	E
White liquor	E	E
Sulfite liquor	E	E
Chlorite bleach	G	F
Alum	E	

PHOTOGRAPHIC INDUSTRY		
Developers	E	G
Silver nitrate	E	G
General use	E	G

FERTILIZER INDUSTRY		
General use	E	E

STEEL INDUSTRY		
Sulfuric acid pickling	E	E
Hydrochloric acid pickling	E	E
H₂SO₄-HNO₃ pickling	F	N

TEXTILE INDUSTRY		
General use	E	E
Hypochlorite bleach	G	F

FOOD INDUSTRY		
General use	E	E
Breweries	E	E
Dairies	E	

MISCELLANEOUS INDUSTRIES		
Plating	G	G
Petroleum	N	N
Tanning	E	E
Oil and soap	F	P
Water and sewer	E	E

RATINGS:

E—No attack
G—Appreciably no attack
F—Some attack but usable in some instances
P—Attacked—not recommended

N—Rapidly attacked
C—Cold—75° F.
H—Hot—125° F.

No. 24

In San Francisco's temperate, yet invigorating climate, pleasantly warm in the daytime, but always cool in the evening, the American Institute of Chemical Engineers will hold a meeting from Sept. 13 to Sept. 16, 1953, with headquarters at the Fairmont Hotel. There will be much to see and do, and participation in the full and interesting program projected, including the presentation of technical papers, plant trips, women's activities, and the various divertissements of the Bay Area will be stimulating and rewarding to chemical engineers and their wives, too.

On the technical side, special interest attaches to the papers on chemical engineering fundamentals and to the symposia on mixing, and on ion exchange. Also from the technical viewpoint, the chemical engineer will welcome the opportunity to see, in the scheduled trips and plant tours, the processes at work in a winery, the processes involved in the making of soap and other detergents, and what goes on inside a radiation laboratory, a well-known center of atomic energy—to mention just a few of the promised sights.

The spirit of San Francisco is symbolized in its two bridges—the gleaming towers of steel that span the Golden Gate and the San Francisco Bay, and in a 400-acre Treasure Island brought forth from the floor of the Bay, which served as the site of the Golden Gate International Exposition during 1939 and 1940, and which now houses a Naval installation.

Throughout the city, visitors will find museums, art galleries, libraries, and theaters. In Lincoln Park, overlooking the Golden Gate, is the Palace of the Legion of Honor, containing one of the outstanding art galleries in the West. At the beautiful marble and granite Opera House in the Civic Center was housed the United Nations Conference on International Organization in 1945, and more recently the Japanese Treaty Conference.

Mention should be made of the unexcelled department stores and shops, the hotels with their varied dining rooms and entertainment, the corner flower stands with their colorful array. All this, and much more spell San Francisco.

Chemical Industry of the Area

CEMENT

The antecedents of the chemical industry in the Bay Area are lost in the colorful tangle of California's early history. All indications, however, point to the cement industry. Benecia Cement Co. built the first plant in 1859, when consumption of cement in San Francisco

This article was prepared by F. G. Sawyer, W. Q. Hull, and Elliot Schrier.

A panoramic view of the city with famous Market Street running down to the Ferry Building. On the right, the San Francisco-Oakland Bay Bridge, Yerba Buena Island and the Companile Tower on the University of California campus across the Bay in Berkeley. On the left, the financial center and Treasure Island.

Let's Go To San Francisco

San Francisco's colorful California Street cable car ascending steep Nob Hill. In the background, San Francisco-Oakland Bridge.

Fisherman's Wharf has all the color and charm of a Mediterranean town.

was only 12,000 bbl. a year. The Benecia product was hydraulic cement, using shell marl (seashells in a matrix of calcite and clay) as raw material.

The area's modern cement industry is largely dependent upon two other natural materials: oyster shells and limestone.

Extensive deposits of limestone are scattered throughout the Coast ranges. Permanente Cement Co. operates one of the world's two largest cement plants, one such deposit near Los Altos, producing more than 5,000,000 bbl. a year.

SALT

San Francisco Bay's unique solar salt industry came into being close on the heels of the cement industry. First the Indians, then the mission fathers, and finally the "Forty-Niners" gathered the salt deposited on the margin of the Bay during centuries of summertime evaporation. The plant near Alvarado at the south end of the Bay, has been in continuous production and is now part of the Leslie Salt Co.

In 1868 seventeen companies with a total capacity of 17,000 tons a year lined the southeastern shore of the Bay. By 1880 the needs of the Comstock Lode had pushed production past 30,000 tons a year. In 1910, the first vacuum-refining unit was installed by Leslie Salt Refining Co. The later history of the industry is largely one of consolidation and mechanization.

MAGNESIA

Development of another industry peculiar to the Bay Area, the large magnesia products and bromine industry, can be traced back to the salt industry. Around 1880 Union Pacific Co., then the biggest in the area, carbonated its

bittern (waste liquor from salt production) with lime kiln gas to make a dynamite absorbent for Hercules Powder Co. Although Union Pacific lasted until 1926, when it was bought and operated as a separate unit by Arden, absorbent production seems to have petered out by the end of the century.

The next significant attempt to utilize bittern came with World War I, when the Bay Area salt companies made potash, magnesium sulfate, and magnesium chloride by fractional crystallization. But the process proved uneconomical in normal markets, and by 1926 the last unit operated by the salt companies was shut down.

In 1923 demand for magnesium chloride became so acute on the West Coast that National Kellastone Co. established the California Chemical Corp. to supply this product. California Chemical's first plant was built in San Diego, but in 1926 operations were shifted to San Mateo. The company gradually expanded its activities and eventually outgrew its parent.

The first United States recovery of magnesium compounds directly from sea water was accomplished by Marine Chemical Co. in 1927 at South San Francisco. Initial production was 1,000 lb. of magnesium carbonate a day. Now the Marine Magnesium Corp., owned by Merck & Co., Inc., makes relatively small quantities of high-priced magnesia products, such as milk of magnesia. Originally, magnesium hydroxide was precipitated with calcined oyster shells, but now calcined dolomite is used.

All the major magnesite producers are located on the Pacific Coast, and most of the caustic-calcined magnesite made in this country is produced in California. One of the early giants in the industry, when oxychloride cement was in its heyday, was Sierra Magnesite Co., founded in 1920 by National Kellastone, then one of the country's largest magnesite consumers. In 1926 high-purity artificial periclase was first produced at Sierra Magnesite's Porterville plant. In 1924 Sierra Magnesite and the California Chemical Corp. were merged

COMMITTEE CHAIRMEN

General chairman .... G. C. Gester
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Technical Program .... R. W. Moulton

to form the California Chemical Co., which was acquired in 1937 by Westvaco Chlorine Products Corp.

EXPLOSIVES

Two years after the first commercial salt works had been established on San Francisco Bay in 1864, the California Powder Works was formed at Santa Cruz to produce the area's first black blasting powder. The company established a dynamite plant at Hercules in 1879, and shifted its Santa Cruz operations there in 1915. The company was taken over first by Du Pont and then by Hercules.

The first United States dynamite was produced by Giant Powder Co. (now Atlas Powder Co.) in 1868 in what is now Sunset District of San Francisco. A year earlier the San Francisco Chemical Works was established in the Mission District to produce nitric and sulfuric acids. The company later also produced nitroglycerin.

PETROLEUM

The first shipment of crude petroleum to reach San Francisco for distilling and sale arrived in 1865, but it was not until 1879 that the first refinery was built. Pacific Coast Oil Co., one of the predecessors of Standard Oil Company of California, operated a plant in Alameda. The oldest refinery still in operation, however, is Union Oil Co.'s plant at Oleum, opened for business in 1879. In 1902 Standard of California placed its Richmond refinery on stream. The Richmond plant is now the largest in the West. Shell and Tidewater-Associated help to swell the tonnage of petroleum shipments.

CAUSTIC-CHLORINE

The Bay Area chemical industry, as known today, received a big push when the Great Western Chemical Co. was formed and ground broken for an electrolytic caustic soda and chlorine industry at Pittsburg. The chlorine was used to make bleaching powder, and the caustic was sold to the oil refineries.

In 1924 Great Western produced the world's first commercial quantities of xanthates. Great Western merged in 1938 with The Dow Chemical Co.

Other points of interest in the Bay Area chemical industry include International Mineral & Chemical Corp. monosodium glutamate plant at San Jose—the only plant of its kind, utilizing waste liquor from beet sugar production; the Owens-Corning Fiberglas plant at Santa Clara; Kaiser Aluminum & Chemical Corp. "chemicals-from-the-sea" plant at Moss Landing; California Spray

Chemical Corp. office and plant (Standard of California subsidiary) at Richmond—home of one of the world's largest manufacturers and distributors of agricultural chemicals, founded in 1903; the main Chlorox plant in Oakland—home of the largest bleach manufacturer in the world; Stanford Research Institute; El Dorado Oil Works; Philadelphia Quartz Co.; Stauffer Chemical Co.; and many others.

The Lighter Vein

Two events have been arranged for the entertainment and participation of all: Get Acquainted Party on Sunday evening in Gold Room of the headquarters Fairmont Hotel, and a typical Mexican luncheon on Monday in the Fairmont's Tonga Room. The luncheon food will be prepared under the supervision of Al Williams, famous restauranteur, whose Papagayo Room at the Fairmont is famous for its authentic Mexican cuisine. During luncheon, the renowned Hurtado Brothers Marimba players will add their artistry to the colorful affair with Mexican and other Latin tunes.

To give out-of-towners freedom to satisfy their culinary tastes and to give them an opportunity to see the Bay Area, no formal dinner events have been arranged. All meeting registrants will receive a list of recommended restaurants as judged by Ray Ewell, gourmet and author of "Dining Out in San Francisco."

The local entertainment committee will arrange for reservations for groups wishing to dine out or to take in typical tourist attractions, such as a cruise around San Francisco Bay, a tour of the city's world famous Chinatown, etc. Information will also be provided on vacation spots and for arranging special trips

before and after the meeting. Those interested in this service and in any of the entertainment features may write J. A. Samaniego, % Shell Development Co., Emeryville, Calif. Other entertainment events of interest are the following: The opening on Sept. 14 of "Kismet," starring Alfred Drake at the Curran Theater; opening of the San Francisco Opera season on the night of Sept. 15 with "Mephistopheles," followed on the 16th with "Othello"; the Palace of the Legion of Honor art feature by early Impressionists; the first U. S. showing of the Rouart Collection; the special showing of South Seas art at De Young Museum.

Ladies' Program

On Sunday evening, Sept. 13, there will be an informal get-together and cocktail party at the Fairmont Hotel. The following day will feature the Mexican luncheon referred to previously. The rest of this day will be available for sightseeing and shopping (shops and stores are open Monday evening).

On Tuesday there will be an all-day trip down the Peninsula. This will include a look at Fisherman's Wharf, the Presidio, Golden Gate Park, Stanford University with special stops at the Stanford Chapel and the Hoover Library. Luncheon will be at the Allied Arts Guild in Menlo Park. Following lunch there will be a tour through the strikingly modern home office of Sunset Magazine. The return to San Francisco will feature some of the fine residential sections of San Mateo and Burlingame. Preregistration of at least one week is advisable for this all-day tour. No formal ladies' program is planned for Wednesday so as to allow additional

time for browsing about San Francisco and environs.

Sartorial Note: The ladies will find a warm topcoat desirable for the evenings. Suits or dresses will be suitable for daytime wear.

Seeing the Sights

Part of the magic of San Francisco lies in its many contrasts. Built on seven hills, the city rises steeply, providing panoramic views of the Bay and the Pacific Ocean. Its cosmopolitan population is reflected in picturesque Chinatown and North Beach and in its famous restaurants. Tall buildings gain additional height from their hilltop locations, and, on the side streets, one sees houses and gardens which literally hang on to the steep slopes.

This can be seen best by swinging aboard one of the cable cars and going up the hills, or walking through Chinatown, which adjoins the city's business and shopping districts. In this transplanted bit of the Orient, are colorful pagodas, intriguing shops, as well as the markets where a busy settlement of Chinese shop for their own needs.

Picturesque San Francisco goes beyond Chinatown and winds slowly through the circling streets of the North Beach area and on to Fisherman's Wharf. Here visitors stroll down sidewalks lined with cauldrons, steaming with the odor of boiling crab and shrimp to the wharves where the little boats of the Italian fishermen are tied up after sailing through the Golden Gate to deposit their day's catch. This same catch comprises the menu in the seafood restaurants surrounding the wharf.

The tour of this section would not be complete unless the visitor went to the top of Telegraph Hill to enjoy the breath-taking view of the city, the Bay, and the great bridges. From this point the view is reminiscent of the romantic and colorful past. At this point on Telegraph Hill signals were transmitted to old sailing ships, giving them instructions to berth their precious cargoes. If a view of the city from the water is preferred, daily boat trips are made around San Francisco Bay—under the Golden Gate Bridge, around Alcatraz Island, Angel Island, and back under the Bay Bridge. A trip by ferry to Oakland and return is easy to arrange as ferries leave every half hour from the Ferry Building, landmark at the foot of Market Street.

A visit to Golden Gate Park, whose thousand acres were reclaimed from sand dunes is another "must." Here can be found the De Young Museum, rare and colorful tropical fish in the Steinhart Aquarium, and the remarkable collection

Richmond refinery, Standard Oil Company of California—largest in the West.

TECHNICAL PROGRAM

Monday, Sept. 14, 1953

TECHNICAL SESSION NO. 1

Symposium on Economic Evaluation of
Chemical Projects

9:30 A.M.—ECONOMIC EVALUATION OF SEVERAL PROCESSES FOR THE PRODUCTION OF AMMONIA SYNTHESIS GAS FROM NATURAL GAS, B. J. Mayland, E. A. Comley and J. C. Reynolds, The Girdler Co., Louisville, Ky.

10:15 A.M.—PLANT COSTS FOR ACETYLENE FROM HYDROCARBONS — COMPARISON OF REGENERATIVE AND PARTIAL COMBUSTION PROCESSES, M. J. P. Bogart and R. H. Dodd, The Lummus Co., New York.

11:00 A.M.—CONSIDERATION OF CAPITAL COSTS FOR A PETROCHEMICAL PROJECT, E. B. Chiswell and J. J. Merrill, California Research Corp., Richmond, Calif.

TECHNICAL SESSION NO. 2

General Technical Program

9:30 A.M.—CO-CURRENT LIQUID-GAS FLOW IN A PIPE-LINE CONTACTOR, G. E. Alves, Du Pont Co., Wilmington, Del.

10:00 A.M.—TWO-PHASE FLUID FLOW, BRIDGING VELOCITIES IN WETTED-WALL COLUMNS, R. R. Kraybill and Brymer Williams, University of Michigan, Ann Arbor, Mich.

10:30 A.M.—A CORRELATION OF FLOODING VELOCITIES IN PACKED COLUMNS, E. H. Hoffing and F. J. Lockhart, The M. W. Kellogg Co., Jersey City, N. J.

11:00 A.M.—THE ROSETTE—A NEW PACKING FOR DIFFUSIONAL OPERATIONS BASED ON HIGH INTERSTITIAL HOLDUP, A. J. Teller, Fenn College, Cleveland, Ohio.

11:30 A.M.—PRESSURE DROP THROUGH POROUS MEDIA, L. E. Brownell D. C. Gami, R. A. Miller, and W. F. Nekarvis, University of Michigan, Ann Arbor.

TECHNICAL SESSION NO. 3

Symposium on Mixing

2:00 P.M.—DISCHARGE FLOW FROM TURBINE TYPE MIXING IMPELLERS, J. P. Sachs and J. H. Rushton, Illinois Institute of Technology, Chicago, Ill.

2:30 P.M.—CONTINUOUS FLOW STIRRED TANK REACTORS. ELEMENTARY CONSIDERATIONS FOR SOLID-LIQUID SYSTEMS, E. L. Piret, R. V. Mattern, and O. Bilous, University of Minnesota, Minneapolis, Minn.

3:00 P.M.—EVALUATING PERFORMANCE CHARACTERISTICS OF MECHANICAL MIXING PROCESSES: THE DEXTROSE - KAOLINITE - WATER

SYSTEM, A. S. Michaels and V. Puzinauskas, Massachusetts Institute of Technology, Cambridge, Mass.

3:30 P.M.—EFFECTS OF AGITATOR GEOMETRY IN LIQUID-LIQUID MIXING, T. Vermeulen, J. L. Fick, and H. E. Rea, University of California, Berkeley, Calif.

4:00 P.M.—THE PUMP-MIX MIXER SETTLER. A NEW LIQUID-LIQUID EXTRACTOR, B. V. Coplan, J. K. Davidson, and E. L. Zebroiski, Knolls Atomic Power Laboratory, Schnectady, N. Y.

TECHNICAL SESSION NO. 4

Symposium on Chemical Engineering
Fundamentals—General

2:00 P.M.—FORMATION OF BUBBLES AT SIMPLE ORIFICES, R. R. Hughes, A. E. Handlos, H. D. Evans, and R. L. Maycock, Shell Development Co., Emeryville, Calif.

2:30 P.M.—THE PRECISION OF MASS SPECTROMETER ANALYSES OF CARBURETTED WATER GAS, W. Volk, Hydrocarbon Research, Inc., Trenton, N. J.

3:00 P.M.—DEVELOPMENT OF A NEW EQUATION OF STATE FOR GASES, J. J. Martin and Yu-Chun Hou, University of Michigan.

3:30 P.M.—LIQUID-VAPOR PHASE BEHAVIOR OF METHANE-ETHANE MIXTURES—PARTS I AND II, O. T. Bloomer, D. C. Gami, I. P. Patel, E. C. Kovacic, and J. D. Parent, Institute of Gas Technology, Chicago, Ill.

Tuesday, Sept. 15, 1953

TECHNICAL SESSION NO. 5

Symposium on Chemical Engineering Fundamentals—Transport Properties

9:30 A.M.—INTERMOLECULAR FORCES AND THEIR PROPERTIES OF GASES, R. B. Bird, J. O. Hirschfelder, and C. F. Curtiss, University of Wisconsin, Madison, Wis.

10:00 A.M.—THERMAL DIFFUSION IN LIQUIDS, H. G. Drickamer and W. M. Rutherford, University of Illinois, Urbana, Ill.

10:30 A.M.—ESTIMATION OF DIFFUSION COEFFICIENTS FOR GASES AND VAPORS, C. R. Wilke and C. Y. Lee, University of California.

11:00 A.M.—DIFFUSION COEFFICIENTS FOR HYDROCARBONS n-HEXANE AND n-HEPTANE IN AIR, W. G. Schlinger, H. H. Reamer, and B. H. Sage, California Institute of Technology, Pasadena, Calif.

TECHNICAL SESSION NO. 6

General Technical Program

9:30 A.M.—SEPARATION OF INDIVIDUAL RARE

EARTHS BY LIQUID-LIQUID EXTRACTION FROM MULTICOMPONENT MONAZITE RARE EARTH NITRATES, J. Bochinski, M. Smutz, and F. H. Spedding, Iowa State College, Ames, Iowa.

10:00 A.M.—STEAM PYROLYSIS OF ETHANE, R. A. Carpenter and F. G. Fowler, Midwest Research Institute, Kansas City, Mo.

10:30 A.M.—HYPERFORMING, Clyde Berg, Union Oil Company of California, Wilmington, Calif.

11:00 A.M.—THE CATALYTIC POLYFORMING OF SHALE OIL, Lloyd Berg, L. G. Mayfield, H. A. Saner, and R. L. Crecelius, Montana State College, Bozeman, Mont.

TECHNICAL SESSION NO. 7

Symposium on Chemical Engineering Fundamentals—Transport Properties

2:00 P.M.—THERMAL CONDUCTIVITY OF GASES AT MODERATE PRESSURES, LeRoy A. Bromley, University of California, Berkeley.

2:30 P.M.—THE THERMAL CONDUCTIVITY OF LIQUID HYDROCARBONS AT LOW TEMPERATURES, R. T. Ellington, Jr., J. D. Parent, and Ralph E. Peck, Institute of Gas Technology, Chicago.

3:00 P.M.—THE VISCOSITY OF GASES AND GAS MIXTURES AT HIGH PRESSURES, N. L. Carr, J. D. Parent, and R. E. Peck, Institute of Gas Technology, Chicago.

3:30 P.M.—THE VISCOSITY OF VAPOR MIXTURES OF HYDROGEN PEROXIDE AND WATER, C. N. Satterfield, R. L. Wentworth, and S. T. Demetriades, Massachusetts Institute of Technology, Cambridge.

TECHNICAL SESSION NO. 8

Symposium on Chemical Engineering Fundamentals—Applied Thermodynamics

2:00 P.M.—PHASE EQUILIBRIA IN SYSTEMS WITH SUPERCRITICAL ETHYLENE, D. B. Todd and J. C. Elgin, Princeton University, Princeton, N. J.

2:30 P.M.—VAPOR-LIQUID PHASE EQUILIBRIA —HYDROGEN CHLORIDE—ETHANE, J. H. Ashley and G. M. Brown, Northwestern Technological Institute, Evanston, Ill.

3:00 P.M.—THE HEATS OF VAPORIZATION OF A BINARY MIXTURE, D. W. Schroeder and W. C. Edmister, Carnegie Institute of Technology, Pittsburgh, Pa.

3:30 P.M.—THE INTEGRAL ISOBARIC HEAT OF VAPORIZATION OF METHANOL-BENZENE MIXTURES, J. A. Tallmadge, D. W. Schroeder, W. C. Edmister, and L. N. Canjar, Carnegie Institute of Technology, Pittsburgh.

A battery of eight miniature stills and a larger still has been installed in Standard Oil's Whiting, Ind., research laboratories for use in exact fractional distillation. The battery uses three types of interchangeable distilling columns. One, the spinning-band column, has been improved by use of a magnetic drive. The band inside the glass column is turned by an outside electric motor that has no physical connection with the band, and thus leakage at the packing glands under reduced pressure has been eliminated. The larger still employs a 1-in.-diam. packed column 16 ft. high, which has a separation efficiency equivalent to more than 300 theoretical plates. The design of the still heads permits fractions as small as 0.02 oz. to be taken off without danger of their being mixed, according to Standard's research department, which uses the stills also in refinery design.

TECHNICAL, SCIENTIFIC REGISTERS PLANNED

Registers of scientific and technical personnel are now being compiled by professional societies with support from the National Science Foundation, which is authorized by Congress to "provide a central clearinghouse for information covering all scientific and technical personnel in the United States." Four societies are already engaged in the program, and among others that have indicated a desire to participate is the Engineers Joint Council.

Up-to-date information on about 100,000 scientists and engineers is expected to be on hand by June, 1954, and by June, 1955, it is expected that registers will have been established in all major fields of science.

The registers, the Foundation states, will become basic tools for supply and demand studies for trained manpower in various fields of science and will serve as stand-by machinery that will aid in the event of an emergency in identifying and locating scientists with special skills.

MONTANA BEGINS PH.D. PROGRAM

The Chemical Engineering and Chemistry departments at Montana State College, Bozeman, were authorized by the Montana Board of Education last month to grant the Ph.D. degree. These will be the first departments in the University of Montana system to offer the doctorate program, which will be started this autumn.

ENGINEERING GRADUATES RECEIVE HIGHER PAY

Average starting salary for the June engineering graduate of Illinois Institute of Technology reached an all-time high of $362 a month, according to a survey conducted by Earl C. Kubicek, placement director. Mr. Kubicek announced that the beginning salary of the June graduate jumped $20 over the February class average of $342. He attributed the rise to the extreme shortage of engineers and to the fact that a large percentage of graduates have previous work experience and are offered premium salaries.

The largest group in the graduating class, and the highest paid engineers, were the mechanical engineers, who averaged $377 a month. Next were the electrical engineers with $371; industrial engineers, $370; mathematics majors, $368; and civil engineers, $367. Slightly below the class average of $362 were physics majors, with $360. Chemists averaged $356; chemical engineers, $352; fire-protection and safety engineers, $342; business and economics graduates, $327; and metallurgical engineers, $317.

"Even those facing military service—60% of the class is nonveteran—were snapped up by industry," Mr. Kubicek said. "Most companies were willing to hire service-eligible engineers in the hope that, even if they were called up after a few months, the experience they gained with a firm and the seniority they accrued while in service would induce them to return to their former employer."

The rise in engineering graduates' monthly salaries, according to Illinois Tech. figures, is illustrated by $355 for 1952; $265 for 1948; $169 for 1943; and $100 for 1938.

A recent survey by the American Chemical Society indicates that, contrary to the Illinois Tech. report, chemical engineers are receiving somewhat higher starting salaries than chemists in 1953. The median salary for chemical engineers, according to the A.C.S. survey, is $360, while that for chemists is $351. A possible explanation is that many chemical engineers are graduates of five-year courses.

TITANIUM PLANT IN GEORGIA FOR CALCO

Calco Chemical Division of American Cyanamid Co. has announced construction of a $14 million titanium dioxide plant near Savannah, Ga., to be started late this year. The plant has been designed by Robert and Company Associates of Atlanta.

A new ammonia plant for Shell Chemical Co., at Ventura, Calif., scheduled for completion this summer. Constructed by The M. W. Kellogg Co., the plant will include a synthesis-gas reformer, which operates at higher than customary pressures and reduces the compressor load required in the later ammonia-synthesis stage. Design capacity is over 150 tons of anhydrous ammonia daily.

NEW PROCESS FOR SULFURIC ACID

The introduction into the United States of a new European process for the manufacture of sulfuric acid by combustion of various sulfur-containing materials such as pyrites, spent oxide, cold gases from copper converters, or gypsum was announced recently by International Processes, Inc., Chicago. Designed to replace the old lead-chamber process, this Kachkaroff-Guareschi Process, according to its sponsors, converts 99.9% of sulfur dioxide entering the system to sulfuric acid. The basis of the process is the reaction, under controlled conditions, of SO_2, O_2, and nitrous vitriol.

Process

In the main system for production of 78% acid, the burner gases pass directly from the dust chamber to a denitration tower and then through the several compartments of the reaction tower. Gases are vented to the atmosphere through a small filter-absorption tower at the end of the plant. The passage of gas to the reaction towers is controlled to make it possible to vary the composition of the liquid in the different towers. The control valves may also be manipulated to change from series to parallel operation in the reactors.

In order to bring the reactants into intimate contact, nitrous acid is circulated over the reaction towers, the heat of reaction being removed by cooling of the circulating acid in coolers. Correct nitrosity is maintained by introducing nitric acid or gaseous products of ammonia oxidation and water to the system. The sulfuric acid so formed in the reaction tower results in an increase in volume of the circulating acid, and this surplus is bled off to the denitration tower. All acid going to the denitrator represents the surplus resulting from volumetric expansion when the acid is formed.

The final product is withdrawn at approximately 78% H_2SO_4, which after cooling is pumped to stock or to a concentration tower, as requirements demand. Traces of acid and small quantities of nitrogen oxides may be carried forward in the gases from the reaction tower, and therefore a small filter absorption tower is installed at the end of the system. Cold sulfuric acid is circulated through this tower in small quantities, the nitrosity being maintained at a constant low value by continuous bleeding to the denitration tower and

Eimco Pilot Filter Stations Give Accurate Data
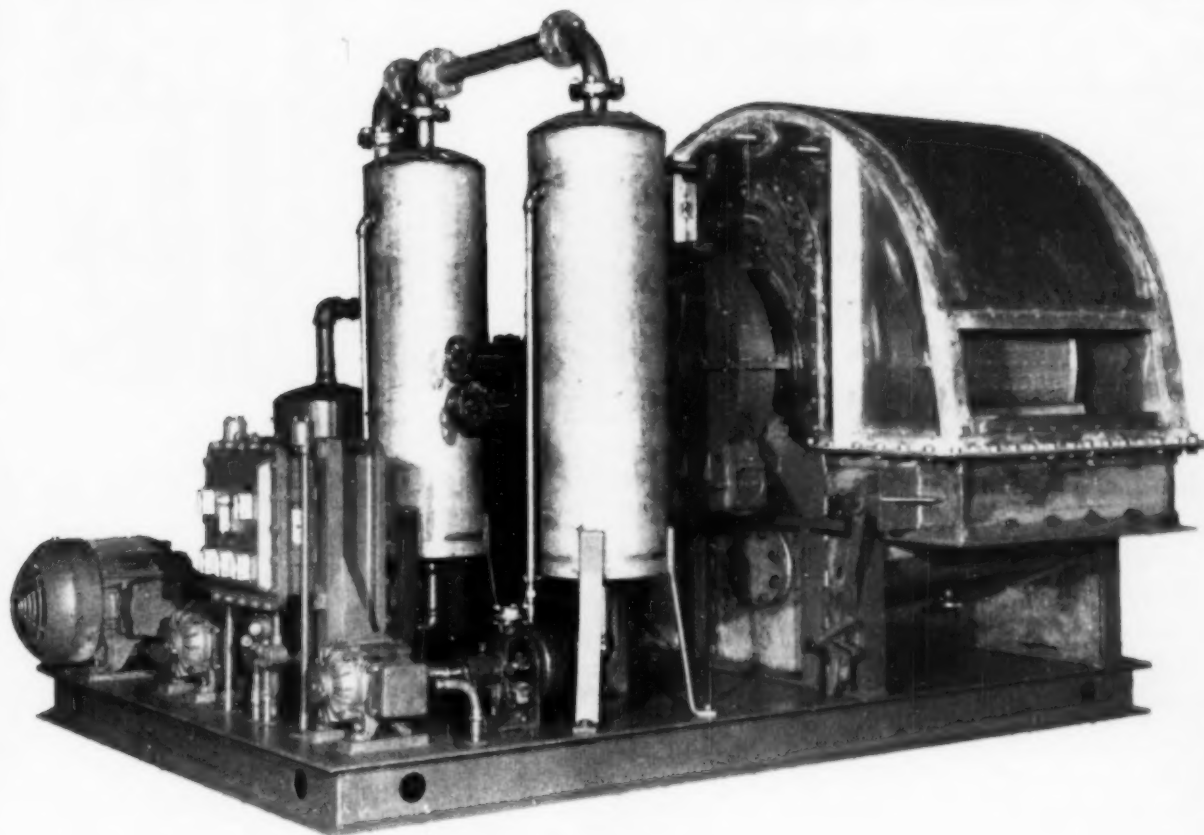
The machine pictured above is another example of Eimco custom filter design and construction. The filter is installed in one of the country's large pharmaceutical plants producing antibiotics and the construction incorporates Eimco's popular package unit design.

The filter unit is a 4 foot diameter by 2 foot face size — of all type 316 stainless steel materials with multiple cake washing apparatus, vapor tight hood, roller discharge and other features. The filter with all of its accessories is mounted on a single platform with all piping and valves connected, all electrical wiring

from motors to a central control panel. The unit arrives at the user's plant ready to run.

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SULFURIC ACID REPORT

replenishment with denitrated acid as required.

Plant

The plant is essentially similar to that used in the lead-chamber system; however the high concentration of all the acid circulated in the system permits the use of iron for a majority of the coolers and pipework. Polyvinyl chloride sheet linings are used extensively. As almost no gaseous nitrogen oxides are liberated, the plants may have a reaction space of as low as 0.4 cu.ft./lb. sulfur/24 hr. One reactor or a bank of coolers may be taken off stream for repairs without shut down of the plant, according to the company. The percentage of sulfur dioxide in burner gases may vary over a period of several hours from 3% or less to 12% without adjustment being necessary.

Process Data

Less than 1 lb. HNO_3/lb. H_2SO_4 and less than 2.3 lb. HNO_3/100 lb. burned sulfur are consumed in this process. The gaseous products of an associated ammonia oxidation plant can also serve as a source of nitrogen oxides. There is less than 0.3 g. SO_3/cu.ft. in the exit gases. Approximately 0.5 cu.ft./lb. of sulfur is produced daily at a power expenditure of 17/20 kw.hr./ton H_2SO_4.

The process may be adapted for the simultaneous production of sulfuric and nitric acid, and a further modification allows for production of sulfuric acid by a small contact plant connected in parallel. As the process allows for the use of cold-sulfur-dioxide gases, advantage may be taken of the heat of the burner gases to concentrate the acid in a tower placed immediately after the dust chamber. By this means, the complete production may be concentrated to about 80% H_2SO_4, or about 40% of the production may be concentrated to approximately 96% H_2SO_4, depending on the temperature of the burner gases.

NATIONAL CYLINDER GAS ACQUIRES GIRDLER CORP.

National Cylinder Gas Co. announced recently the acquisition of The Girdler Corporation of Louisville, Ky., which will be known as The Girdler Company, a Division of National Cylinder Gas Company. Through the merger National Cylinder Gas has acquired the stock of Tube Turns, Inc., of Louisville and Pennsylvania Forge Corporation of Philadelphia, both of which will continue as separate companies. Girdler will continue to operate from Louisville with little change in its organization.

MATHOLIN FORMED FOR HYDRAZINE PRODUCTION
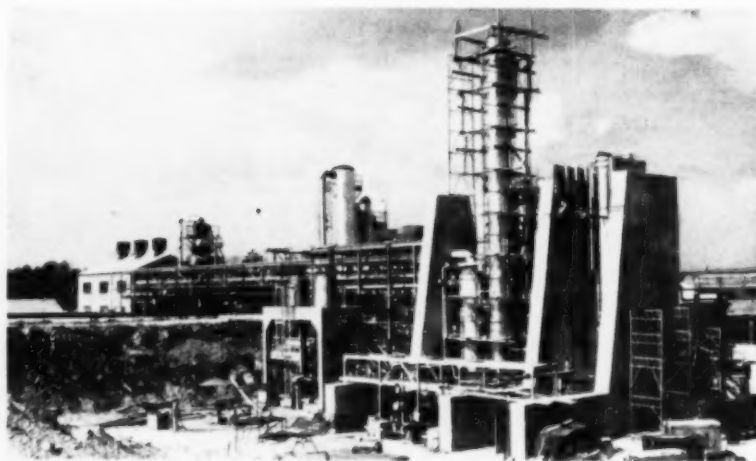
Matholin Corp. hydrazine plant at Lake Charles, La. The distillation unit is seen in the foreground.

Formation of the Matholin Corporation, to be devoted to the production of hydrazine, has been announced by Mathieson Chemical Corp. and Olin Industries, Inc. The new company acquired the first commercial-scale hydrazine-producing unit in this country, recently completed by the Mathieson Corp. at Lake Charles, La.

Belonging to the hydronitrogens, a group analogous to the hydrocarbons, hydrazine is a clear, colorless liquid closely resembling water. Its density and boiling point are close to those of water. Hydrazine, however, is one of the most highly reactive of the inorganic chemicals and is a powerful and concentrated source of chemical energy. Heretofore the best known member of the hydronitrogen family was ammonia.

Modern hydrazine production is based on the Raschig method of synthesis, which involves the reaction of ammonia with sodium hypochlorite. As developed by Mathieson engineers, the process has resulted in anhydrous hydrazine, considerably superior to the earlier hydrate, which is 36% water. Manufacture of anhydrous hydrazine (95% hydrazine) is achieved through ammonolysis or azeotropic distillation.

Besides anhydrous hydrazine and the hydrate, the Lake Charles plant will also produce monohydrazine sulfate, dihydrazine sulfate, hydrazine hydrobromide, and hydrazine hydrochloride. Most of the initial output is expected to be required for military use, but industrial possibilities are being developed. The principal military interest in hydrazine is in its high useful-energy release, particularly from the anhydrous form, which offers potentialities as rocket fuel and for other explosive uses. It is in this field that Olin Industries, which manufactures arms, explosives, and ammunition, is prominent. Industrial uses of hydrazine result from its strong deoxidizing, or reducing, action. It may be used as a reducing agent to separate rare metals from their oxides, to plate thin coatings of metals on glass, and to remove dissolved oxygen from boiler water, thus preventing corrosion. Hydrazine may also be used as a foaming agent in the manufacture of foam rubber.

Hydrazine, which has been known since its discovery in Germany about 75 years ago, did not come into prominence until the Germans used it for rocket fuel during World War II. Mixed with an oxidant such as hydrogen peroxide or liquid oxygen, hydrazine as a fuel enabled German planes to climb at the rate of 7 miles/min. The two German plants, which operated from 1944 to 1945, carried the process through to hydrazine hydrate only. In 1948 its military importance and industrial potentialities led the Mathieson Corp. to build a pilot plant for the manufacture of hydrazine after nearly ten years of research, sponsored in part by the Government.

When still a laboratory product, hydrazine sold for $50 a pound, and by 1949 pilot plant quantities had reduced the price to $9. Today the price is $2.50 a pound, on a hydrazine basis, for the hydrate in solution. Further process improvements may eventually bring the price down to about 50 cents a pound, according to the Mathieson Corp.

Mathieson and Olin equally own the Matholin Corp. John M. Olin, president of Olin Industries, is the president of the new company, and the board of directors is comprised of four representatives each from the parent companies.

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MARGINAL NOTES

A Condensation of Much Experience

Volumetric and Phase Behavior of Oil Field Hydrocarbon Systems. M. B. Standing, Reinhold Publishing Corp., New York (1952), 123 pp. (69 figures, 35 tables), $10.00.

Reviewed by K. H. Hachmuth, Research Planning Board, Phillips Petroleum Co., Bartlesville, Okla.

This book is aimed directly at oil field technologists and as such is highly specialized and not likely to interest many outside the fields of petroleum production and petroleum reservoir engineering. Few engineers in other fields would be justified in adding it to their permanent library although the book is well worth leafing through for the ideas and suggestions that may help on their own special problems.

For engineers about to start a career and for those already engaged in petroleum production, the book is excellent. It condenses in one volume the background and special knowledge necessary for a capable approach to the solution of oil production problems as affected by volumetric and phase behavior of hydrocarbons. Others experienced in this field may not always agree with Dr. Standing's approach but they must admit he proceeds directly to the point without including a miscellany of confusing alternatives or conjectures. The book covers a remarkable amount of ground for its length.

The first chapter concisely reviews the scientific concepts involved. The second chapter describes sampling methods and apparatus, discusses precautions to be observed and tells how to check validity of samples. This chapter alone can save the beginning engineer mistakes that he would avoid at almost any cost. The next three chapters discuss the behavior of gases, of liquids and vapor-liquid equilibria with specific attention to oil field hydrocarbon systems. The two types of petroleum reservoir systems producing liquid hydrocarbons, the gas-condensate and the dissolved gas systems, are discussed in the next two chapters. The final chapter, fittingly, covers the general principles involved.

The type used and the large pages, 8½ by 11 in., make the book easy reading. Its many charts, including three folded charts, give data of special utility in solving oil production problems involving volumetric and phase behavior. The use of these charts is illustrated by an ample number of example problems

MARGINAL NOTES

and often by more than one way of solving a given problem. Tricks of the trade are revealed, for instance, a method for revising available equilibrium data to make it consistent with observed results.

Statistics and the Cuckoo

The Methods of Statistics, 4th ed., L. H. C. Tippett, John Wiley & Sons, New York (1952), 395 pp. $6.00.

Reviewed by Cuthbert Daniel, Engineering Statistician, New York.

The fourth edition of Tippett's text is an interesting and even charming book. With characteristic modesty, the author writes in his preface, and reiterates later, that he has only gone as far in each branch of the subject as his own knowledge and experience permit. Expanded more than a hundred pages from its predecessor, this edition includes many original teaching devices. The visualization of randomization on page 55 is one of the best of these. But most of the familiar examples from the first edition are still there. Perhaps the serious engineer of 1953 will not be as intrigued as this reviewer was in 1931 with the statistical analysis of the connection between the lengths of cuckoo's eggs and the species of the host-birds in whose nests the eggs were laid. Many of Tippett's examples, however, are from textile research, and several from other areas of engineering interest.

The general plan of the book follows the usual order. After a chapter on frequency distributions and their parameters, comes one on distributions derived from the theory of probability. Chapters VI and VII are on the analysis of variance, regarded by the author as the most useful of the tools of modern statistics. The next four chapters are on correlation and "regression," a statistical term referring here mainly to curve fitting.

There are many things to praise in this book, but the engineer with a beginning interest in statistical methods will want to know if it can be recommended to him as a starter. The answer can only be yes to the engineer, who is ready to commit himself to several supplementary texts since the topic judged above to be of major value is not adequately handled, even at the elementary level. Since a similar qualification would have to be made for any other single introductory text, perhaps this judgment will not be considered too adverse. Several important matters are

CANDIDATES FOR MEMBERSHIP IN A. I. Ch. E.

The following is a list of candidates for the designated grades of membership in A.I.Ch.E. recommended for election by the Committee on Admissions.

These names are listed in accordance with Article III, Section 7, of the Constitution of A.I.Ch.E.

Objections to the election of any of these candidates from Active Members will receive careful consideration if received before September 15, 1953, at the Office of the Secretary, A.I.Ch.E., 120 East 41st, New York 17.

Applicants for Active Membership

Ackerman, Carl D., *Pittsburgh, Pa.*

Baird, William George, Jr., *Lockport, N. Y.*

Benson, Homer E., *Pittsburgh, Pa.*

Breidenbach, Eric P., *Baton Rouge, La.*

Brown, William W., *Ambler, Pa.*

Butenhoff, Melvin W., *Port Arthur, Tex.*

Crowley, B. A., *Baltimore, Md.*

Darling, Carlton L., *Wickliffe, Ohio*

Dunlap, Ralph I., *Springfield, Mass.*

Englander, Alan J., *Palmetto, Fla.*

Fisher, Frank R., *New York, N. Y.*

Frame, John M., *Cincinnati, Ohio*

Gilbert, Carlisle W., *Pittsburgh, Pa.*

Gore, James, III, *Coraopolis, Pa.*

Greever, James E., *Cincinnati, Ohio*

Hays, George E., *Bartlesville, Okla.*

Hogeland, A. William, *Ridley Park, Pa.*

Jansen, J. J., *North Wilbraham, Mass.*

Kraftson, Harry A., *Bala-Cynwyd, Pa.*

Krasnoff, Albert, *Maplewood, N. J.*

Laster, Richard, *Forest Hills, L. I., N. Y.*

Laundrie, Robert W., *Akron, Ohio*

LeMay, J. E., *Dallas, Tex.*

Ludwig, Raymond N., *Port Arthur, Tex.*

Massey, Lester G., *Chicago, Ill.*

McCants, Malcolm T., *Corpus Christi, Tex.*

McQuarrie, J. Graham, *Texas City, Tex.*

Mills, Donald N., *Houston, Tex.*

Myers, Charles E., *Bon Air, Va.*

O'Connell, Francis P., *Metuchen, N. J.*

Osborn, Oliver, *Lake Jackson, Tex.*

Osri, Stanley M., *Glenview, Ill.*

Paddison, O. H., *Tenafly, N. J.*

Pollock, J. A., *Akron, Ohio*

Rice, Theodore, *Pittsburgh, Pa.*

Rice, William J., *Houston, Tex.*

Rich, R. E., *Notre Dame, Ind.*

Robertson, James M., *Kingsville, Tex.*

Ross, S. David, *Drexel Hill, Pa.*

Russell, Henry H., *Pittsburgh, Pa.*

Schmidlein, Joseph A., *Darien, Conn.*

Schwartz, George N., *St. Louis, Mo.*

Serdarian, Sarkis, *Wilbraham, Mass.*

Shurter, Robert A., *Terre Haute, Ind.*

Stansel, J. C., *Coshocton, Ohio*

Swanson, Paul F., *Short Hills, N. J.*

Weedman, John A., *Bartlesville, Okla.*

Wilcox, Allan C., *Midland, Mich.*

Willis, James M., *Akron, Ohio*

Winn, Hugh, *Akron, Ohio*

Wood, J. Q., *Bartlesville, Okla.*

Yeakley, Vernon J., *New York, N. Y.*

Zahnstecher, Leonard W., *New York, N. Y.*

Zeiger, H. Evan, *Shawmut, Ala.*

Applicants for Associate Membership

Bucherer, Walter F., *Montreal, P.Q., Canada*

Dorsheimer, Wesley T., *Stamford, Conn.*

Francis, Gus, *Denver, Colo.*

Porter, H. K., *New Castle, Del.*

Rosenfield, Myer J., *Cincinnati, Ohio*

Applicants for Junior Membership

Ameen, Raymond F., *Pawtucket, R. I.*

Anderson, Malcolm M., *Charleston, W. Va.*

Andrew, Eugene A., Jr., *St. Louis, Mo.*

Ball, Frank J., *Springfield, Mass.*

Baylor, H. Nelson, *Charleston, W. Va.*

Baxley, William H., *Hartsville, S. C.*

Bazell, Clyde, *Ironton, Ohio*

Beals, Richard N., *Miamisburg, Ohio*

Benes, Peter, *Cleveland, Ohio*

Bennett, Thomas M., Brooklyn, N. Y.

Bierman, Robert L., Cincinnati, Ohio

Blaney, William H., Lake Jackson, Tex.

Blumberg, Sheldon R., University Hts., Ohio

Bosshardt, Robert C., Terre Haute, Ind.

Bowers, Donald L., Logan, Ohio

Bradshaw, Robert Daily, Gibsonburg, Ohio

Briggs, Roger L., Corning, Ohio

Brinson, Kenneth, Terre Haute, Ind.

Broulette, James J., Seattle, Wash.

Burkhart, Robert L., Mansfield, Ohio

Bussolari, Robert J., Wilmington, Del.

Byars, Harry G., Fort Worth, Tex.

Carl, Ralph W., Boonville, Mo.

Carmichael, Charles J., Jr., Houston, Tex.

Cartier, Raymond Martin, New York, N. Y.

Cashman, John M., Arsenal, Ark.

Casto, William H., Charleston, W. Va.

Chambers, William C., Anderson, Ind.

Chen, Alice Lee, New York, N. Y.

Christensen, Donald E., Tacoma, Wash.

Cinque, Jack J., Texas City, Tex.

Clarkson, W. Philip, Suffern, N. Y.

Close, John C., Highland Park, Ill.

Coe, Benjamin P., Schenectady, N. Y.

Condra, Louis W., Jr., Lake Jackson, Tex.

Conklin, Charles White, Jr., Smithfield, Va.

Crossley, Walter C., Brownsville, Tex.

Curth, Douglas R., Louisville, Ky.

Dagley, Leo J., Jr., Bartlesville, Okla.

Davis, Akin T., Washington, D. C.

Dean, George W., III, Hopewell, Va.

DeGisi, Sabino Lucas, Louisville, Ky.

Delacretaz, Richard E., Springfield, Mass.

DeLand, Donald L., Barberton, Ohio

Dienes, George L., New York, N. Y.

DiGiacomo, Angelo A., Phila., Pa.

DiPietro, George A., Elsmere, Del.

Doherty, Bernard E., Toledo, Ohio

Dohn, Leo A., Jamaica, N. Y.

Donham, Walter E., Whiting, Ind.

Donnelly, James R., Denver, Colo.

Dorsey, A. Gorman, Jr., Baltimore, Md.

Dresher, William H., Phila., Pa.

Duke, Donald A., Texas City, Tex.

Duncan, Robert W., Bordentown, N. J.

Dunmyer, Joseph C., Jr., Freedom, Pa.

Edeleanu, Alexander G., Richmond, Calif.

Egerter, George, Philadelphia, Pa.

Eglington, Arthur R., Newark, Del.

Eisen, Stephen, New York, N. Y.

Elenniss, Chris C., Lakewood, Ohio

Emmerling, Fred A., Rochester, N. Y.

Emmett, Robert, Jr., Chihuahua, Mexico

Ernst, Richard T., Louisville, Ky.

Felsen, Harold, Hunter, N. Y.

Findlay, Donald E., Columbus, Ohio

Fisher, John A., Fostoria, Ohio

Fleck, Brice E., So. Charleston, W. Va.

Fleischer, Edward G., Bloomfield, N. J.

Fletcher, Kenneth J., Wood-Ridge, N. J.

Francais, Richard K., Kokomo, Ind.

Freistedt, Robert R., Skokie, Ill.

Frew, Charles P., Charleston, W. Va.

Fusco, Joseph J., Drexel Hill, Pa.

Galbreath, Richmond B., New Orleans, La.

Garrison, Lawrence J., New York, N. Y.

Gatewood, Herbert L., Terre Haute, Ind.

Gough, Robert R., Hopewell, Va.

Gay, Philip T., Tewksbury, Mass.

George, Wendell J., Detroit, Mich.

Gingrich, Rey W., Toronto, Ontario, Can.

Gipe, David A., Jr., Johnstown, Pa.

Goizueta, Roberto C., Havana, Cuba

Goldman, Edward, New Hyde Park, N. Y.

Grubaugh, Richard F., Indianapolis, Ind.

Haltom, Ed. O., Jr., Dallas, Tex.

Hannah, James F., Columbia, S. C.

Harding, Ronald H., W. Lafayette, Ind.

Hershey, Daniel, Brooklyn, N. Y.

Hertel, Paul H., Philadelphia, Pa.

Hewitt, Robert T., Jr., Columbus, Ohio

Higgins, Robert D., Montreal, Quebec, Can.

Hiller, Alfred E., New Orleans, La.

Hoffman, Raymond, Brooklyn, N. Y.

Holt, Vell D., Chattanooga, Tenn.

Holtzapple, John, Jr., Wayne, Pa.

Holzman, Allen L., Belleville, N. J.

Howard, John F., Wilmington, Del.

Hunt, Marshal, Jr., Hammond, Ind.

Huntington, Richard L., Findlay, Ohio

Jaros, Robert J., Baltimore, Md.

Jurczak, Henry Edward, Chicago, Ill.

Kahlson, Don, Cleveland, Ohio

Kalvinskas, John J., Woodbury Heights, N. J.

Kaplan, David L., Cincinnati, Ohio

Kauffield, Robert C., Conneaut, Ohio

Kirchner, Carl E., Jr., Lakewood, Ohio

Kisciras, Paul G., Rutherford, N. J.

Kladko, Martin, Seneca Falls, N. Y.

Klaus, R. Alan, Terre Haute, Ind.

Klein, Arthur J., Jr., Freeport, Tex.

Klein, Donald J., Taunton, Mass.

Knapp, Todd, LaCrosse, Wisc.

Knudsen, Irving E., Rosedale, N. Y.

Koch, George E., Lorain, Ohio

Koeninger, Edward C., Cincinnati, Ohio

Kohler, Richard B., Forest Hills, L. I., N. Y.

Korn, Alfred E., Belleville, N. J.

Koskela, Otto A., Texas City, Tex.

Kowal, Richard F., Akron, Ohio

Krieger, James H., Cleveland, Ohio

Kueteman, Herman H., Texas City, Tex.

Kumler, P. Richard, Lancaster, Ohio

LaBerge, John C., Baltimore, Md.

Lauerty, Albert George, Paterson, N. J.

Lawrence, William J., Phila., Pa.

LeCompte, S. Walter, Jr., Easton, Md.

Lehmann, W. Kemp, Baltimore, Md.

Lewis, Julian B., Wauwatosa, Wisc.

Link, Walter K., Collingswood, N. J.

Lipscomb, Walter P., So. Charleston, W. Va.

Littman, Jack, Upton, N. Y.

Llewellyn, George H., Jr., Columbia, S. C.

Lombardi, Crispino, Plainfield, N. J.

Loter, William, Paterson, N. J.

Louis, Howard, Marcus Hook, Pa.

Lubitz, Herbert H., New York, N. Y.

MacGregor, Donald, II, Rochester, N. Y.

Mahoney, John G., Youngstown, Ohio

Markert, George B., Lancaster, Pa.

Maurice, Roger L., Charleston, W. Va.

McKee, David B., Granite City, Ill.

In

Large

Refineries

and Small...
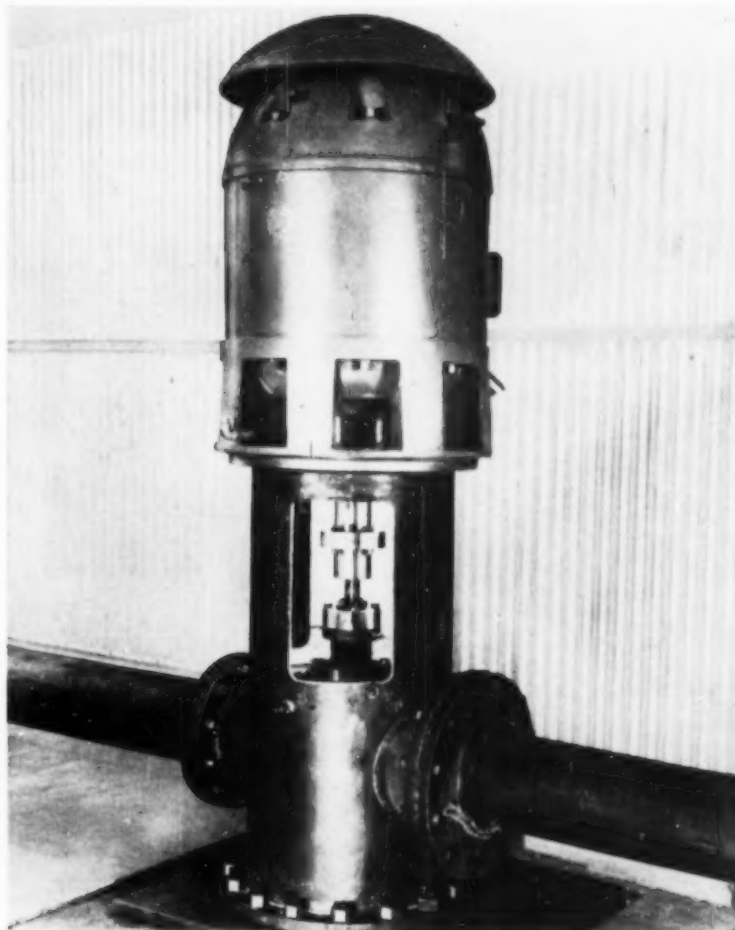
JOHNSTON PUMPS ARE DOING TOUGH JOBS HANDLING VOLATILE FLUIDS

Here's a Johnston, installed in a large Southern California refinery, where it is used as a high-octane gasoline transfer pump. This unit, which has the pumping element enclosed in a sealed steel barrel, takes suction sufficiently above the pump bowls to give the required net positive suction head. The discharge is at the same level as the suction. Such a unit offers many advantages:

1. **The supply vessel can be located wherever convenient and the pump bowl is then set low enough for proper submergence.**
2. **Driver and piping connections are at the** surface, eliminating dry sumps which are **costly to build and expensive to maintain.**
3. **Nothing to lubricate but driver bearings.**
4. **Vertical design needs only ⅓ as much valuable floor space.**
5. **Due to diffuser type construction, this unit has higher efficiencies than conventional centrifugal pumps at high heads.**

This is another typical example of the ability of Johnston equipment to handle all types of pumping jobs. Why not let our trained engineering staff consult with you on the best solution to YOUR particular problem?

Send today for your free copy of the new bulletin "Vertical Pumps for Volatile Fluids." Ask for Bulletin G-83.

JOHNSTON PUMP COMPANY
3272 EAST FOOTHILL BLVD., PASADENA 8, CALIFORNIA

PRODUCTS

1FC   Dryers
Custom-designed and tested dryers.
Louisville Drying Machinery Unit, General American
Transportation Corp.

3R   Varidrive Motors
Any R P M by simple dial control. Varidrive accomplishes in a single power unit the functions of a
motor, speed changer, gear box and controls, to
occupy little more space than a fixed-speed motor.
Bulletin.
U. S. Electrical Motors Inc.

4A   New Centrifugal Pumps
Impervious graphite equipment for processing, conveying, and storing corrosive fluids. Bulletins.
National Carbon Co.

6L   Filtration Equipment
Heavy-duty filter. 60-sec. opening or closing. Also
standard horizontal plate models.
Sparkler Manufacturing Co.

7A   Durco Valves
Corrosion resisting valves with Teflon Sleeves, available
in ¼ in. through 2 in. flanged and screwed.
The Duriron Co.

8L   Tank Gauging
Varec remote reading tank gauge receivers, equipped
with 36 key-type switches. Receiver design is flexible
permitting custom panel installations when desired.
Bulletin.
The Vapor Recovery Systems Co.

9A   Acetylene
Wulff process plants to produce acetylene at low cost
high-purity acetylene from readily available hydrocarbons such as natural gas, propane, and butane.
Ethylene can be produced as a by-product.
Girdler Co.

10A   Tel-O-Set Controllers
Fast-acting pneumatic controls. Sealed-tight design, non-
bleed pilot valve, simplified tuning, easily-read dial
calibrations. Applications cover pressure, liquid level,
flow and temperature. Specification Sheets 768 and
771.
Minneapolis-Honeywell Regulator Co.

12L   Photochemical Equipment
For continuous flow, batch or recycling processes. Literture.
Hanovia Chemical & Mfg. Co.

13A   Precessing Kettles
Available in research laboratory to test new processes
and new products. Kettles perform a number of basic
operations; heating, cooling, mixing, extracting, reacting, distilling, evaporating, drying, and solvent recovery. Vacuum, atmospheric, or pressure operations
can be provided. 48-page color booklet.
Buflovak Equipment Division.

14A   Heat-Transfer Medium
Dowtherm for chemical, paint, petroleum, food and
other process industries.
The Dow Chemical Co.

15A   Compressors
Air or gas, for pulling vacuums. Capacities from 5 to
125 hp.
Ingersoll-Rand

16A   Proportioning Pump
Capacity range from 1 to 10 gal./hr. will discharge
against pressures up to 650 lb./sq.in. gauge. Bulletin 1105.
Proportioneers, Inc.

392A   Packing Tower
Intalox saddle packing for distillation towers.
U. S. Stoneware

21A   Gasholder
Dry seal permanent, impermeable and weatherproof gas-
holders.
Wiggins Gasholder Div., General American Transportation Corp.

23A   Acetylene from Hydrocarbons
Manufacturing acetylene economically by the Wulff
Process.
The Fluor Corp., Ltd.

24L   Metering and Proportioning Pumps
U type pumps for small volume flows, standard, jacketed,
and variable speed units. Catalog UP-52R.
Hills-McCanna Co.

25A   Pilot Filter Stations
This filter unit is a 4-ft. diam. by 2 ft. face size—of all
type 316 stainless steel materials with multiple-cake-
washing apparatus, vapor-tight hood, and roller discharge.
The Eimco Corp.

27A   Controlled Volume Pump
For hydrochloric acid . . . all concentrations chlorinated hydrocarbons, corrosive metallic salt solutions,
dilute acids, and mixed acids. Write for Bulletin 300.
Lapp Insulator Co., Inc.

28L   Controlled Humidity
Hygroll moisture-absorbent liquid removes moisture as
a separate function from cooling and heating.
Niagara Blower Co.

29A   Graphite Anodes
Chlorine and caustic soda for the production of many
plastic materials.
Electrode Div., Great Lakes Carbon Corp.

30L   Rotary Dryers
Hersey preheaters, dryers, kilns, coolers and calciners.
Bulletin No. 531.
Standard Steel Corp.

31A   Celite Filtration
Provides pure process water for boiler make-up, heat
exchangers, and circulating water for towers.
Johns-Manville

32L   Demineralizing
New technique of mixed-bed demineralization.
Barnstead Still & Sterilizer Co.

33A   Stainless Steel Valves
Cast stainless steel valves, fittings and accessories. 2-in.
valve comparison chart.
The Cooper Alloy Foundry Co.

34B   Plug Valves
Nonlubricated valves open and close with a quarter-
turn.
DeZurik Shower Co.

35A   Volatile Fluid Pump
Transfer pumps for volatile fluids, high octane gasoline,
etc. Bulletin C-83.
Johnston Pump Co.

Chemical Engineering Progress

Numbers followed by letters indicate advertisements, the number corresponding to the page carrying the ad. This is for ease in making an inquiry as you read the advertisements. Letters indicate position—L, left; R, right; T, top; B, bottom; BL, bottom left; BR, bottom right. A indicates a full page; IFC, IBC, and OBC are cover advertisements.

Be sure to give name, address, position, etc.

Remember, the numbers on the upper portion of the card bring you data on only the bulletins, equipment, services, and chemicals reported in these information insert pages. The lower portion of the card is for the advertised products, and is keyed not only to advertising pages, but also to the memory-tickling list under the heading Products.

36A   **Control Valves**
Examples of Hammel-Dahl control valves in various chemical plants.
Hammel-Dahl Co.

43A   **Turbogrid Distillation Trays**
A new type distillation tray for replacement of bubble-cap installations.
Shell Development Co.

47A   **Dorrco Fluosolids System**
Produces SO_2 for contact acid plant and desulfurized zinc calcine for electrolytic zinc production.
The Dorr Co.

48T   **Filter Presses**
Filters for any material. Catalog.
T. Shriver & Co., Inc.

48B   **Steam Traps**
Nicholson thermostatic steam traps for process, heat, power applications. Catalog.
W. H. Nicholson & Co.

49A   **Glass Pipe**
Corrosion-resistant and noncontaminating Pyrex pipe. Manual.
Corning Glass Works

50L   **Industrial Equipment**
Dust and fume control, spray booths, mechanical washers, and industrial ovens.
Schmieg Industries Inc.

50R   **Molybdenum Trioxide**
Raw material for the production of all molybdenum compounds, including molybdenum orange, molybdenum vermilion, and molybdated pigment colors.
Climax Molybdenum Co.

PRODUCTS (Continued)

51A **Chilled Water**
Hermetic centrifugal refrigeration units, 45 to 200 tons, for either comfort or process.
The Trane Co.

52T **Spray Nozzles**
Full-cone, flat- and hollow-cone nozzles.
Spray Engineering Co.

52B **Filter Presses**
A complete range of filter bases and a choice of six different closing devices. Catalog.
D. R. Sperry & Co.

53A **Stainless Tubing**
For heat-exchanger applications. Chart.
The Carpenter Steel Co.

54L **Pumps**
Pumps for boiler feeding, heavy-duty compressor for general use, and duplex compressor for larger plants and soot blowing. Catalog and bulletins.
Pennsylvania Pump & Compressor Co.

55A **Fintube Heaters**
Tank suction heaters and line heaters. Bulletin No. 521.
Brown Fintube Co.

57A **Tank Insulation**
Zerolite insulation for outdoor tanks and vessels.
Johns-Manvillle

58B **Deionizer**
Single-tank, mixed-bed ion-exchange unit produces de-ionized water for less than 25 cents/1000 gal. Bulletin 512. Water softeners: Bulletin 611.
Elgin Softener Corp.

59A **Diaphragm Valves**
Packless valves in brass, steel, iron.
Crane Co.

60T **Poro-Stone Filters**
A package plant for special and standard filtering problems. Model for continuous filtration molten phosphorus. Bulletin No. 431.
R. P. Adams Co., Inc.

60BL **Spun Trays**
Continuous, vertical, transfer type for sensitive food products. Special bulletin on spun trays for dryers, coolers and toasters in the food industry.
Wyssmont Co.

60BR **Technical Data Books**
Data books contain about 140 pages of technical data, essential for the student, engineer, technical worker, and business man.
Lefax

61R **Tantalum**
Acid-proof tantalum equipment for chemical operations. Booklet.
Fansteel Metallurgical Corp.

62L **Chemical Exposition**
In Philadelphia the first time Nov. 30 to Dec. 5. Application forms for hotel accommodations available.
International Exposition Co.

63R **Screening Material**
Wire cloth woven from all commercially used metals in sizes ranging from 4 in. (space cloth) to 400 mesh. 5,400 different combinations. Catalog E.
Newark Wire Cloth Co.

64L **Color Comparators**
For pH, chlorine, phosphate, and nitrate tests. Free handbook.
W. A. Taylor & Co.

65T **Furane Resin Cement**
A mortar for corrosion-resistant masonry. Folder.
Delrac Corp.

65B **Proportioning Pumps**
7,500, 15,000 or 30,000 lb./sq.in. working pressure by the interchange of piston and cylinder assemblies. Bulletins 4061-A.
American Instrument Co., Inc.

66T **Bubble Caps**
Bulletin 21 contains specification information for more than 200 standard styles of bubble caps and risers. Special caps designed.
The Pressed Steel Co.

66B **Pumps**
Synthetic rubber products handled by Nagle pumps. Also used for abrasive and corrosive applications.
Nagle Pumps, Inc.

67L **Heat Exchangers**
Paracoil heat exchangers, for pilot-plant units, and plant modernization programs.
Davis Engineering Corp.

67R **Tricone Mills**
Bulletin AH-414-40 for ball mill information.
Hardinge Co., Inc.

69T **Laboratory Ware**
Vitreous silica transparent, glazed and unglazed crucibles, evaporating dishes, beakers, tubing. Bulletins.
The Thermal Syndicate, Ltd.

69B **Agitator Drives**
Drive reduction units of all descriptions. Booklet No. 5308.
Western Gear Works

71R **Antifoam**
Silicone defoamer in recirculating cooling brines, bottling soft drinks, varnish cooking, textile resin backings, and steel pickling baths. Coupon for free sample.
Dow Corning Corp.

72T **Viscometer**
Adaptable to any problem from less than one to 32,000,000 centipoises.
Brookfield Engineering Laboratories, Inc.

72B **Plate Fabrication**
Towers, pressure vessels and heat exchangers. Literature.
Downingtown Iron Works, Inc.

73T **Tube Rolling**
Expand tubes to produce uniformly tight tube joints. Bulletin TX-50.
Vernon Tool Co., Ltd.

73B **Evactors**
Chill-Vactors chill water in plants cooling chemical solutions, fruit juices, milk, whiskey mash, etc.
Croll-Reynolds Co., Inc.

IBC **Heat Exchangers**
For process conditions ranging from liquid air to Dowtherm, high vacuum to high pressure.
The Vulcan Copper & Supply Co.

OBC **Mixers**
Units for a variety of fluid mixing operations. Coupon for catalogs.
Mixing Equipment Co., Inc.

CHEMICALS

Industrial Chemicals. Atlas Powder Co. catalog on seven classes industrial chemicals. Sorbitol, surfactants, mannitol, solid polyester resins, fatty acid, plasticizers, and activated carbons.

1

Pelargonic Acid. Improved grade of pelargonic acid in experimental and pilot quantities. Technical bulletin lists specifications, characteristics, reactions, applications. Emery Industries, Inc.

2

Dispersing Agents. Dewey and Almy Chemical Co. announce potassium salt dispersing agent in powder form for latex paint industry. Low sodium and sulfate ion content control efflorescence.

3

Flattening Agent. Flattening Agent for alkyd-urea varnishes from Davison Chemical Corp. Gives matte, hand-rubbed effect, easily ground. Leaflet.

4

Trioxane. Crystalline form of formaldehyde Trioxane in commercial quantities from Celanese Corp. Stabilizer for trichloroethylene and zein solutions, setting agent for protein materials, etc. Bulletin.

5

Aerosil. A white finely divided silica called Aerosil. Developed by Godfrey L. Cabot, Inc. 99% pure silicon dioxide. Used by rubber compounders in white and colored rubber products.

6

Ammonium Thiocyanate. J. T. Baker Chemical Co. bulletin on ammonium thiocyanate. Gives properties, aquous solutions, uses in various industries. Section on chemical reactivity and use in synthesis.

7

Fatty Alcohols. Revised data sheet from M. Michel and Co., Inc. on Cachalot fatty alcohols. Typical analyses for fourteen types.

8

Polyvinyl Chloride. Ampcoflex rigid polyvinyl chloride by Atlas Mineral Products Co. Bulletin on resistance characteristics, physical properties, techniques of un-plasticized fabrication.

9

pH Catalog. Handbook catalog from LaMotte Chemical Products Co. on equipment and chemical reagents. For determining pH, chlorine control, water, sewage analysis, soil testing, etc.

10

Sodium. Brochure on handling of sodium from Ethyl Corp. Discusses bulk sodium, sodium sand, handling dispersed sodium, applications, etc.

11

Lacquer Emulsions. Hercules Powder Co. brochure on emulsifying agents, and emulsifying and formulation techniques.

12

Teflon and Rulon. Folder on several grades of Teflon, also Rulon from Dixon Saddle Co. Teflon as parts and also as rods, tubing, shapes. Rulon bearings require no lubrication. Withstand temperature $-100°$ F. to more than $500°$ F.

14

Pliovic Resins. Goodyear Tire & Rubber Co. folder on Pliovic polyvinyl chloride. Approximate intrinsic viscosity 0.80. Complements use of grade previously announced for calendering, coating, and extruding.

15

Plastisols. Bulletin on Hysol series of plastisols from Houghton Laboratories, Inc. Corrosion- and abrasion-resistant.

16

Fertilizer Laws. Data book of facts and figures for fertilizer manufacturers from Monsanto Chemical Co. Gives state laws for fertilizer manufacture and sale, charts, formultions, etc.

17

Zinc Dust. Applications of zinc dust described in bulletin from American Smelting and Refining Co. Advantages of use in bleaches, reducing agents, catalyst purifiier, additive to lubricants and paints.

18

Resistant Coatings. New coatings of Lithcote Corp. For corrosion and contamination resistance to many chemical products and exposures. Linings suitable for tank cars, storage tanks, other processing equipment.

19

Plasticizer. Paraplex, polymeric vinyl plasticizer, by Rohm & Haas Co. Properties of interest to formulators of vinyl products. Good resistance to migration in rubbing varnish, auto lacquer, Plexiglas, polystyrene.

20

Alcowax. Allied Chemical & Dye Corp. announces Alcowax derived from ethylene. Melting point $100°$ C. Hard, tasteless, white translucent color. Used in paper coating, polishes, candles, paint.

21

Polyester Resins. General Electric Co. technical bulletin on industrial applications for polyester resins. Discusses catalyst, fillers, pigments used with resins. Also fabricating methods. Numerous tables and charts.

22

Kel-F Polymers. M. W. Kellogg Co. bulletin on physical, electrical, chemical, and mechanical properties of tri-fluorochloroethylene polymers. Information on resistance to more than 100 specific chemical substances.

23

Tygon Paint. Data on Tygon paint, and Tygorust vinyl primer from U. S. Stoneware Co. Gives chemical-resistance charts; tells where and how to use materials.

24

Asphalt Products. Pioneer Latex & Chemical Co., Inc. bulletin on latex, rubber, and asphalt products. Included are asphalt cut-backs, outdoor protective coatings, industrial floors, expansion joint compounds, etc.

25

Amyl Acetate. Data sheet on amyl acetate from Commercial Solvents Corp. General information, specifications, uses, physical properties included.

26

Magnesia. Booklet on General Magnesite & Magnesia Co. products. Types, uses, analysis, material details.

27

Metallic Phosphates. "Application of Metallic Phosphates to Cooling Water Treatment," a paper from Deady Chemical Co. Flow chart of system, data on analyses of water used in tests.

28

Ferrocyanides. Bound volume on chemistry of ferrocyanides from American Cyanamid Co. Summarizes known physical and chemical properties of sodium and potassium ferrocyanides and hydroferrocyanic acid. Covers applications.

29

Rubber-Base Coating. Chemical- and corrosion-resistant rubber-base coating by Saran Protective Coatings Co. Formulated with copolymer of styrene and butadiene. For chemical-food-processing plants, oil refineries, etc. Coatings air-dry, dust-free quickly.

30

Plasticizer. Monsanto Chemical Co. plasticizer compatible with many synthetic and cellulose resins; Santicizer 160. Technical bulletin available. Low volatility at extrusion and calendering temperatures, good oil extraction, heat and light stability.

31

Hydraulic Fluid. Lindol HF, a tricresyl phosphate-based flame-resistant hydraulic fluid announced by Celanese Corp. Physical properties and specifications.

32

Plastic-Coating Process. Process announced by Pyramid Plastics, Inc. for application of plastic coating on flexible steel tubing, metal- and paper-jacketed tubing, etc.

34

Fumaric Acid. Commercial quantities of fumaric acid available from Monsanto Chemical Co. Crystalline, free-flowing granule, material is companion to phthalic and maleic anhydrides. Used for surface coatings, inks, etc. — **35**

Controlling Corrosion. For use in control of corrosion in steam-condensate systems Hagafilm by Hagan Corp. Description, advantages, etc. in leaflet. Improves valve and pump operation, keep lines clear; reduces trap maintenance. — **36**

Wulff Process Acetylene. Lummus Co. technical and economic study on Wulff Process for production of acetylene from light hydrocarbons. Process description, simplified flow diagram, extensive data. — **37**

Furnace Black. Continex fast extrusion furnace black by Witco Chemical Co. discussed in extensive bulletin. Covers properties compared with other blacks, strength, modulus, hardness, other features. — **38**

Niacin - Niacinamide. Reilly Tar and Chemical Co. now producing U. S. P. niacin and niacinamide for use in vitamin preparation. — **39**

Modified Polystyrenes. Data on Koppers Co., Inc. modified polystyrenes included in binder of technical bulletins. Applications, types, properties, manufacture, etc. — **40**

Hot Melt Adhesive. Announced by National Adhesives vinyl type hot melt adhesive for binding. Data on tests comparing grip, toughness, etc. — **41**

Anti-Seize Compound. Lead-plate, an antiseize compound from Armite Laboratories saves maintenance costs by prevention of high-temperature freeze, leakage, corrosion, galling and seizing in all threads to metal-to-metal contacts. Temperatures $-100°$ F. to $+2987°$ F. Effective to pressures of 6000 lb./sq. in. — **42**

BULLETINS

Stainless Bars and Wire. (50) Data on application of stainless wire. Physical properties, corrosion resistance, selection chart. (51) First edition on stainless bars. — **50**

Tables on sizes and shapes, weights, corrosion resistance. Fabrication and processing information. Allegheny Ludlam Steel Corp. — **51**

Gas and Diesel Engines. General bulletin on gas, Diesel, and gas-Diesel engines by Cooper-Bessemer Corp. Atmospheric, supercharged, vertical 4 cycle, 6, 7, 8 cylinders. 690 to 2500 hp. — **52**

Control Literature. Bulletin, Bristol Co., offers check list of available literature on recording and control instruments. — **53**

Cracking Unit. Bulletin on new model fluid catalytic cracking unit from Standard Oil Development Co. No throttling slide valves, lower height, improved vessel design, simpler internals; flow diagram. — **54**

Process Pumps. Ingersoll-Rand pumps for the refinery and process industries. Medium duty services. Sizes 1 to 3 in.; capacities 20 to 550 gal./min.; pressures to 300 lb., temperatures to 500° F. — **55**

Temperature Controls. From Burling Instrument Co. Illustrated catalog of electric temperature controls. New models, selection and comparison table. Some suitable for 1800° F. — **56**

Multi-Point Recorder. Foxboro Co. information sheet on control and alarm combination with multi-record recorder. Optional control and alarm functions. Diagrams clarify each of four control or alarm actions. — **57**

Gravimetric Feeder. New Omega Machine Co. limited space, loss-in-weight feeder for low rates of feed. Delivers material to ¾ in. size; 0 to 100% settings; minimum 0.25 lb./hr.; maximum 100 lb./hr.; contains a capacity selector. — **58**

Wood Sole Shoes. For protection of those in large variety of industries, line of wooden sole shoes from F. J. Stahmer Shoe Co. Catalog illustrates various available types. — **59**

Bucket Elevators. Specification tables, illustrations, schematic drawings of bucket elevators and buckets by Jeffrey Mfg. Co. Catalog. — **60**

Research and Engineering Service. Brochure from Commonwealth Engineering Co. of Ohio discusses service from inception of idea to fulfillment for a wide area of technical interest. — **61**

EQUIPMENT

Space Filter. For air purification an ultra-aire space filter from Mine Safety Appliances Co. Protects against inert and radioactive dusts. Bulletin gives engineering details, diagrams, etc. — **130**

Industrial Thermometers. Bulletin from American Machine and Metals, Inc. on a bronze case industrial thermometer. Three types of connections. Information on scale ranges, frost insulation, shapes, other specifications. — **131**

Solids Feeder. New Richardson Scale Co. solids feeder. Delivers constant flow of material to process equipment at 3000 lb./min. Handles feeds, limestones, dry, free-flowing chemicals, etc. — **132**

Combustion Furnaces. Hevi Duty Electric Co. new organic combustion tube furnace. Three separate furnaces mounted on rails for use as one unit or three individual units, each with temperature control. Sizes, weights, other specifications listed. — **133**

Pressure Transmitter. Moore Products Co. announces a new pneumatic pressure transmitter. Bulletin gives schematic diagram, range chart, element material. No geared movement. — **134**

Sodium Peroxide Bombs. Parr Instrument Co. sodium peroxide bombs for combustion reactions with sodium peroxide. Four sizes for macro, semimicro and micro tests. Bulletin. — **135**

Strain Indicator. Portable, self-contained unit for field and laboratory measurement of strain in structures and machines. Measuring range 0 to 2000 microinches/inch. Baldwin-Lima-Hamilton Corp. — **136**

Safety Grating. Globe Co. safety grip-strut grating. Diamond-shaped pattern, open area over 75%. Struts joined by integral saddles. Sheet steel or aluminum. — **137**

Non-Skid Multiwall Bag. Application of special coating on face and back of bag develops skid resistance, affording protection against breakage, reduces damage in transit. Does not obscure printing. Union Bag & Paper Corp. — **138**

Feeding Device. For introducing chemicals into high-pressure vessels, feeding device by American Defibrator, Inc. Primarily for feeding wood chips and cellulose materials. Sizes 15 to 21 in. in diam. 70 to 100 tons of wood chips/day. — **139**

Signal Advance!
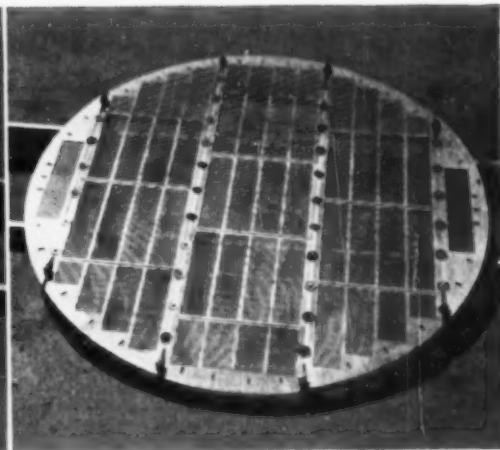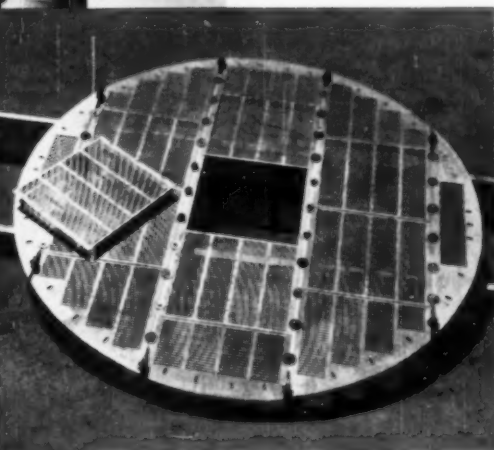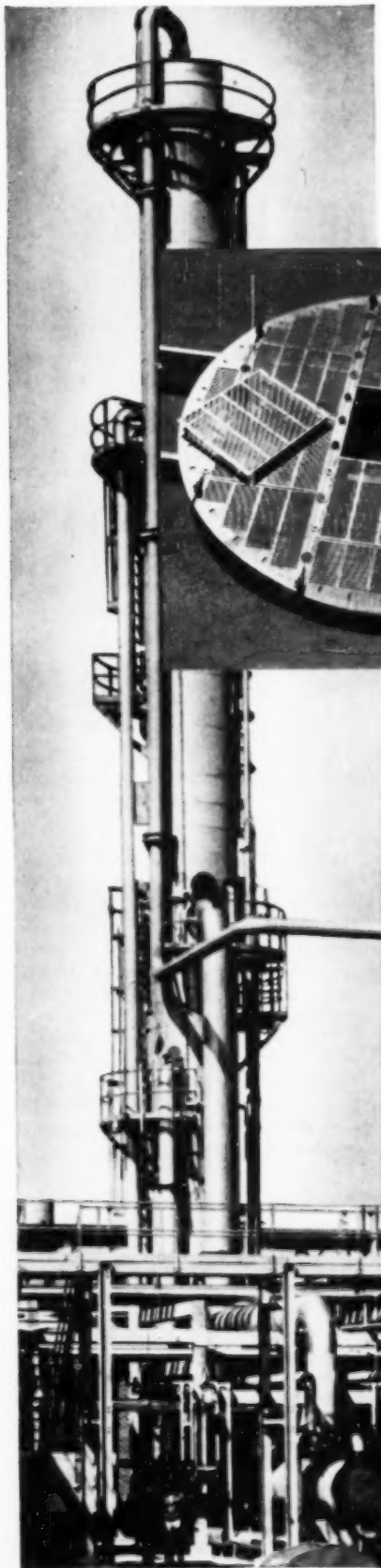
Photos courtesy Fritz W. Glitsch & Sons, Inc.

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the distillation tray that makes the bubble-cap tray obsolete

50 Turbogrid columns in service and as many more under construction prove that—

Turbogrids save money and really deliver
Turbogrid trays save as much as 50% of the cost of *new* column construction.

At nominal cost, tray *replacement* with Turbogrids may even double the capacity of an existing column . . . without loss in product separation.

Ease of control and lower maintenance costs pay additional dividends.

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Visit us at the San Francisco AIChE meeting in September.

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NOMINATIONS ARE IN ORDER

For the past several years local sections of the Institute have been encouraged to suggest suitable candidates for the nomination ballot for A.I.Ch.E. Director. This year to acquaint the members with local section choices, "C.E.P." presents these three pages of candidate biography. Ballots from the Secretary's office are now in the mail, and we hope these sketches help our members in selecting directors for the election ballot. If this service proves to be valuable, "C.E.P." will make it a yearly feature.

AKRON

R. P. Dinsmore, vice-president, The Goodyear Tire & Rubber Co., has been suggested by the Akron Section as A.I.Ch.E. Director. Since joining Goodyear in 1914, he has devoted his entire professional life to that organization, serving in many capacities—on the technical staff, as assistant chemist in Canada, and chief chemist in California and Ohio, organizing the synthetic rubber research and development program for the government and his company. Subsequently, he became vice-president and director, The Goodyear Synthetic Rubber Corp. He was also nominated by the El Dorado, Rocky Mountain, and Boston local sections. He was councilor, The Ohio State University Research Foundation (1951).

ATLANTA

The director of Georgia Tech's Engineering Experiment Station since 1952, Herschel H. Cudd, has been suggested by the Atlanta Section. With a broad background of industrial experience, stemming from his position with Du Pont in the rayon technical division, from directing the inorganic chemical research for International Minerals and Chemical Corp., and from his position as director of the Lantuck division of West Point Manufacturing Co., Dr. Cudd was appointed head of the chemical sciences division, Georgia Tech Engineering Experiment Station in 1950. Much of the success of last year's A.I.Ch.E. meeting in Atlanta is attributable to his work as general chairman.

BATON ROUGE

The chairman of the Committee responsible for designing the A.I.Ch.E. Questionnaire, Lloyd B. Smith, now engaged in administrative work in the Esso Laboratories, is the choice of the Baton Rouge Section. Early in his career he was associated with the Atlantic Refining Co. and in 1939 joined the Standard Oil Development Co., on the staff of the Esso Laboratories, Standard Oil Company of Louisiana, Baton Rouge, to have supervision of chemical pilot plants. Active in the A.I.Ch.E. for more than twenty years, he was one of the original members of the Philadelphia-Wilmington Section and one of the organizers of the Baton Rouge Section of which he is past-chairman. New Orleans Section also suggested Smith, who is a registered professional engineer in Louisiana.

CENTRAL OHIO

The Central Ohio candidate is Frank C. Croxton, technical director, Battelle Memorial Institute, Columbus, Ohio, since 1947. A Ph.D. from the Ohio State University in 1930, he served as a research chemist with the Standard Oil Company of Indiana until 1939 when he joined Battelle. He has served A.I.Ch.E. in several capacities: as first chairman, reorganized Central Ohio Section in 1948 and as a member, Local Sections Committee in 1949; co-chairman, Forty-third Annual Meeting; and on Membership Committee in 1951. He is on the board of trustees, Denison University Research Foundation. As technical director of Battelle, he guides its program of chemical and chemical engineering research.

CHICAGO

W. L. Faith, the suggested nominee of the Chicago Section, has been director of engineering, chemical division, Corn Products Refining Co., Argo, Ill., since 1948. Dr. Faith's educational and professional career covers many years of teaching. He has been special research assistant in chemical engineering, Engineering Experiment Station, University of Illinois; assistant professor, chemical engineering, Kansas State College; professor, chemical engineering, State University of Iowa. He was on leave with Office of Production Research and Development, W.P.B., Washington, D. C., as consultant, assistant director, and deputy director (1942-1945). He has served on many A.I.Ch.E. committees, such as Student Chapter, Program, Admissions.

CLEVELAND

Curry E. Ford, a practicing engineer since 1934 and a recent appointee to the position of manager, Carbon Sales, Union Carbide and Carbon Corp., has been nominated by the Cleveland Section, which he was instrumental in creating, serving as its chairman in 1947 and working on its Executive Council for ten years. With respect to the national organization, he is now serving as chairman of the Local Sections Committee and previously was a member of this and the Membership Committee. His experience in convention work is broad. He was vice-chairman of the 1944 and 1952 Cleveland meetings and was a member of the Planning Committee of the 1948 meeting in that city.

COASTAL GEORGIA

Arthur R. Bookout, Jr. research supervisor, Hercules Powder Co., Brunswick, Ga., has been picked by the Coastal Georgia Section. With Hercules since 1943 at Radford, Va., Experiment Station and home office in Wilmington, Del., and at the Brunswick plant, he has conducted research and pilot-plant development in explosives, distillation, high-pressure synthesis, etc. Prior to his association with Hercules, Bookout served as textile chemist with Kerr Bleaching and Finishing Works, Concord, N. C., and as chemist with the U. S. War Department in Radford, Va. Having helped to organize the Coastal Georgia Section, he has served as member of its Executive Committee, as vice-chairman, and chairman.

COLUMBIA VALLEY

W. Kelly Woods, manager, advance technology, engineering department, Hanford atomic products operation, General Electric Co., Richland, Wash., is the choice of the Columbia Valley Section. A graduate of Massachusetts Institute of Technology with a D.Sc. in chemical engineering, he later taught chemical engineering there and then became associated with the Du Pont Co. in the engineering department. In 1946 he was made head, pile engineering, technical section, Hanford Works, General Electric Co., subsequently becoming manager, pile technology, and then assistant manager, technical section, Hanford Works. Dr. Woods kept the Hanford reactors in operation appreciably beyond their anticipated life.

DETROIT

The chairman, department of chemical and metallurgical engineering, University of Michigan, and consulting engineer, Donald L. Katz, is the selectee of the Detroit Section. For three years prior to his association with the University of Michigan where he has been assistant, associate, full professor, chemical engineering, he filled the position of research engineer, Phillips Petroleum Co., Bartlesville, Okla. In Institute affairs he has been active on such committees as the Program, Nuclear Energy, Education and Accrediting, and Publication. He won the Hanlon Award, Natural Gasoline Association of America, 1950.

MARYLAND

The Maryland Section has come through with the name of M. C. Molstad, director, department of chemical engineering, University of Pennsylvania, Philadelphia, and a consulting chemical engineer. Early in his career Dr. Molstad was associated with the Fixed Nitrogen Research Laboratory, Department of Agriculture, as chemical engineer. Subsequently he went to Yale University as instructor, becoming later an assistant professor. For two years he worked in the ammonia department, Du Pont, and then in 1939 went to the University of Pennsylvania first as associate professor and later became professor of chemical engineering. He has served on many committees of the Institute, including Student Chapters, Admissions, Program, and from 1950 to date, as chairman of the Sub-Committee on Institute Lecture.

NEW YORK

L. P. Scoville, the choice of the New York Section, after a short time with Southwest Engineering Co., entered the service of The Texas Co. and was responsible for the chemical engineering design of many large refinery process installations. With the formation of the Jefferson Chemical Co. in 1944, he became chief engineer of the new concern and is now vice-president in charge of production. Scoville has held a number of offices in the New York Section, A.I.Ch.E., including the chairmanship. In the national organization he has served on the Committee on Admissions and is currently chairman of the Program Committee and a member of the Publications Committee and the Publications Board. He was also named by New Jersey Section.

NORTHEASTERN NEW YORK

A. W. Davison, director of research, Owens-Corning Fiberglas Corp. from 1949 and trustee of the Denison Research Foundation since 1943, has been suggested by the Northeastern New York Section. A B.S. from Denison (1910), an A.M., Ohio State (1911), a Ph.D., Cornell in physical chemistry (1914) and a D.Eng. from Rensselaer (1942), Dr. Davison worked as a chemistry assistant at Ohio State and later at Cornell. From 1921-1925 he was professor of physical chemistry, Rensselaer Polytechnic Institute and from 1925 to 1942 served as the William Weightman Walker professor of chemical engineering and head of the department. In 1943 he assumed the position of scientific director, research laboratory, Owens-Corning.

OHIO VALLEY

The manager of Vulcan Engineering Division, Vulcan Copper and Supply Company of Cincinnati, Ohio, Raphael Katzen, has been selected by the Ohio Valley Section. In his present position he is responsible for direction of a large group of process project mechanical and development engineers. He has been associated with the Northwood Chemical Co. as chemical supervisor and chemical engineer and was its Research Fellow at Brooklyn Polytechnic Institute (1940-1942). He also served the Diamond Alkali Co. as technical supervisor. Dr. Katzen assisted in the formation of the Cleveland and Ohio Valley Sections and is a member of several National committees of A.I.Ch.E.

PHILA.-WILMINGTON

Philadelphia-Wilmington Section has suggested G. E. Holbrook, now assistant director, development department, Du Pont Co., Wilmington, Del., where he started in 1933 as a research chemist. He is also the choice of New Jersey Section. For several months during 1952 he was on loan to the government, serving as director, chemical division, N.P.A. Both locally and nationally, he has worked for the objectives of A.I.Ch.E.—locally, on the Executive Committee, vice-chairman and chairman, Philadelphia-Wilmington Section; nationally, as present chairman of the Committee on Future of the Institute, a member of the Constitution and By-Laws Committee, and a Director of the Institute (1950-1952).

PITTSBURGH

C. C. Monrad has been selected as candidate by the Pittsburgh Section. Currently he is Columbia-Southern research professor of chemical engineering and head of the department, Carnegie Institute of Technology. Dr. Monrad has served as chemical engineer, Standard Oil Company (Ind.); assistant group executive, petroleum unit, National Defense Advisory Commission; consultant on chemical industry, W.P.B.; chief, butadiene and styrene branch, Office of Rubber Director; and consultant, Reconstruction Finance Corp. He has been active on committees, Pittsburgh Section; member, Patents Committee, Program, Student Chapters, and others of A.I.Ch.E.

ST. LOUIS

John J. Healy, Jr., who has a long record of active service in the Institute in both local and national affairs, is the selectee of the St. Louis Section. Associated with Monsanto Chemical Co. in various capacities both in Boston and St. Louis since 1921, he has been assistant to the vice-president in charge of research, development and patents since 1951. From 1948 to 1950, when he served as a Director of the Institute, he was also chairman of the Local Sections Committee. He served as general chairman of the Swampscott meeting in 1950 and is currently technical program chairman for the annual meeting to be held in St. Louis in December. He is also one of A.I.Ch.E. representatives on the Engineering Manpower Commission, Engineers Joint Council. Healy is a graduate of Harvard (A.B.) and M.I.T. (B.S.).

SOUTH TEXAS

Both the South Texas and the Texas Panhandle Section chose William A. Cunningham, professor, chemical engineering, University of Texas. He served as department chairman (1942-1943, 1947-1949) and will serve again for the biennium (1953-1955). His primary technical interests are in the fields of industrial waste disposal and in mineral utilization. Bill Cunningham has grown up professionally with the chemical and petroleum refining industries of the Southwest. He received his M.S. in chemical engineering from the University of Texas and was engaged in industrial work for eight years before returning for part-time work which led to his Ph.D. in 1941. He has been an active member of the South Texas Section, serving on the Executive Committee and as program chairman, and chairman of the Section. Nationally, he has served on the Professional Legislation Committee and currently is a member of the Industrial Waste Disposal Committee.

TERRE HAUTE

He is professor of chemical engineering, Purdue University and at present the administrative head of the Point Four project for engineering education in Formosa. His name is R. Norris Shreve, the suggested nominee of the Terre Haute Section. His industrial experience covers assignments with Mallinckrodt Chemical Co., Lamar Chemical Works, Shreve Chemical Co., Orth and Hastings Corp., and Calco Chemical Co. Professor Shreve went to Purdue in 1930 and served as professor of chemical engineering and, as head of the chemical and metallurgical engineering school from 1947 to 1951. He investigated the chemical industry of Europe in 1928, Japan in 1952, and Formosa in 1953.

TWIN CITIES

Jose B. Calva, the choice of the Twin Cities Section, has been active in chemical engineering for more than thirty years. He heads his own group of research engineers, the J. B. Calva Co., Minneapolis, Minn. Back in 1929 he joined Baker-Adamson and a year later went to St. Paul to work for Minnesota Mining & Manufacturing Co. Mr. Calva helped form the Twin Cities Section of A.I.Ch.E. and has served as its vice-chairman and chairman. He was recently elected president, Minnesota Society of Professional Engineers, Minneapolis Chapter. He obtained a B.S. in chemical engineering at the University of Pennsylvania and the University of Mexico.

WESTERN MASSACHUSETTS

This New England Section suggested Allan W. Low, division engineer, plastics division, Monsanto Chemical Co., Springfield, Mass., as nominee. With Monsanto he has held the positions of general superintendent, technical superintendent, and assistant division engineer, division engineer, plastics division. He served the Allied Chemical & Dye Corp., New York, as production supervisor, research chemical engineer, assistant manager, development division; the West Virginia Ordnance Works, General Chemical Defense Corp., and as technical supervisor, General Chemical Co. A graduate of the University of New Hampshire, Mr. Low has always been interested in activities of the Western Massachusetts Section. He was a member of its Program Committee (1950-1951).

WESTERN NEW YORK

Robert Burns MacMullin, who in 1946 established a consulting practice in the field of chemical and electrochemical engineering and is senior partner of the firm, R. B. MacMullin Associates, Niagara Falls, N. Y., has been chosen by the Western New York Section. For twenty-five years he was employed by the Mathieson Alkali Works, Inc. (Mathieson Chemical Corp.) serving successively as research chemist, engineer, assistant director of research, and manager of development. He has served on the Publications Committee of the Institute for five years, and in 1939 was chairman of the Western New York Section. During the last war his activities were concerned chiefly with the development of a new process for making metallic magnesium and the building of the Lake Charles, La., magnesium plant for the Defense Plants Corp.

Alcan First to Roast Zinc Concentrates with DORRCO FLUOSOLIDS System
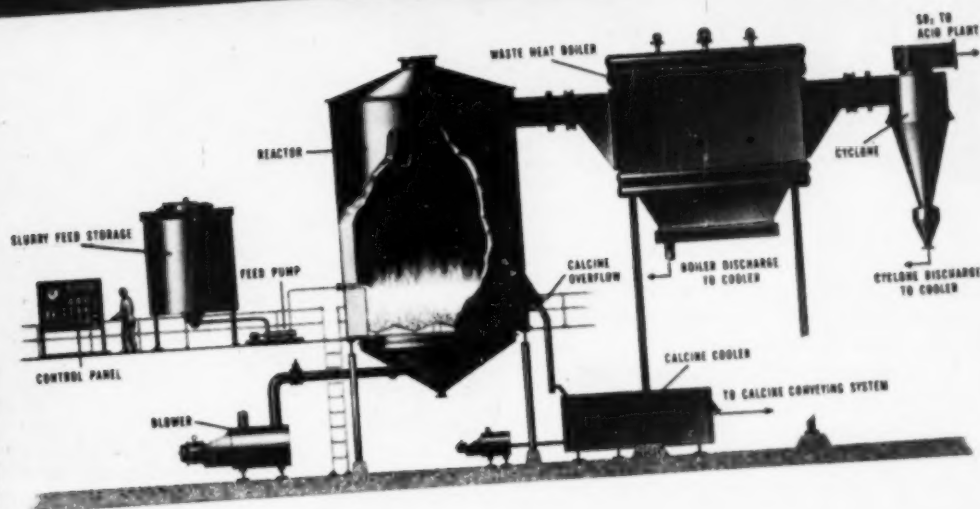
Produces SO₂ for Contact Acid Plant and Desulfurized Zinc Calcine for Leaching

ARVIDA, QUEBEC. First commercial installation for fluidized zinc roasting, a Dorrco FluoSolids System has been on stream at the Aluminum Company of Canada's plant here since June, 1952. Alcan's expanding primary aluminum capacity has increased their sulfuric acid requirements and FluoSolids was selected as the most economically and technically feasible method to produce SO_2 from zinc sulfides for contact acid manufacture. It frees this company from fluctuations in natural sulfur supplies and at the same time gives a calcine ideally suited for electrolytic zinc production.

SYSTEM ROASTS 150 TONS PER DAY

Results from the first year of operation have been completely satisfactory. Gas strength at the Reactor stack averages 10-12% SO_2 . . . a safe margin over the 7% SO_2 required at the converter. The close temperature control possible with FluoSolids has resulted in a combined calcine product averaging 0.3% sulfide sulfur . . . which is the specification on this calcine for electrolytic zinc.

HOW THE SYSTEM WORKS

Zinc sulfide concentrates, obtained from the Northern Quebec areas, are repulped and held in slurry form at about 80% solids in Dorr Agitators. Slurry is pumped into the Reactor and is immediately dispersed and brought to the uniform roaster temperature of 1600°F. Gas passes out the side of the Reactor, through a waste heat boiler and cyclones before going to the acid plant. Calcine is cooled and stored prior to shipment to the zinc refinery.

FOR ZINC PRODUCERS . . .

FluoSolids represents a development of major significance. As proven by the Alcan installation, with this new technique a low uniform temperature can be maintained to produce a calcine with high zinc solubility . . . and at the same time good waste heat recovery can be effected — nearly nine-tenths of a pound of steam per pound of zinc concentrate . . . a strong gas is produced averaging 10-12% SO_2 . . . and there is no necessity for drying nor for fine grinding the roaster feed. In those cases where calcine is sintered, any desired amount of sulfur can be retained, still producing a high strength gas. If you'd like more information on FluoSolids . . . the most significant advance in roasting technique in the last 30 years . . . write The Dorr Company, Stamford, Conn., or in Canada, The Dorr Company, 26 St. Clair Avenue East, Toronto 5.

FluoSolids is a trademark of the The Dorr Company, Reg. U. S. Pat. Off.

MARGINAL NOTES

better treated in this text than in most of those known to the reviewer.

Among the serious defects of the book are: its calculation of significance levels from data on pages 147, 177, 186, 194, 196 (significance levels have to be set independently of the data); its failure to give and use the distinctions between Model I, Model II and Mixed Model experiments; the incorrect definition and then omission of factorial experiments (The experience of K. A. Brownlee and of many others including the reviewer, places these multifactor designs among the most useful in laboratory, pilot plant, and plant experimentation).

Cinquante Ans de Perfectionnement Technique. Presses Documentaires, Paris, France (1952). 380 pp. 4.120 frs port compris.

This volume containing thirty or more articles, represents fifty years of the progress in methods and techniques employed in various industries throughout France. The techniques were described at thirty-two meetings during 1950 before the Centre de Perfectionnement Technique, Paris. The many contributors to this work, all specialists in their respective fields, tell a story which, with its graphs and photographs, gives a complete picture of the advances made in diverse industries, such as metallurgical, perfumery, plastics, oil, lubrication, ceramics, and organic.

Chemical Investments—An Analysis of Research, Growth and Finances in the Chemical Industry. John F. Bohmfalk, Jr., Chemonomics, Inc., New York (1952), 93 pp. $5.00.

The purpose of this work is to interpret with facts and figures the relation of research and sales growth for major end-product groups in the chemical industry. It examines twenty companies in the light of these relationships, discusses the management and research activity, and analyzes the outlook for these individual companies.

Books Received

ASTM Standards on Paint, Varnish, Lacquer, and Related Products. American Society for Testing Materials, Philadelphia 3, Pa. (September, 1952). 800 pp. $5.75.

Reducing Phenol Wastes from Coke Plants. Prepared by Steel Industry Action Committee. Ohio River Valley Water Sanitation Commission (1953), 36 pp. $1.00.
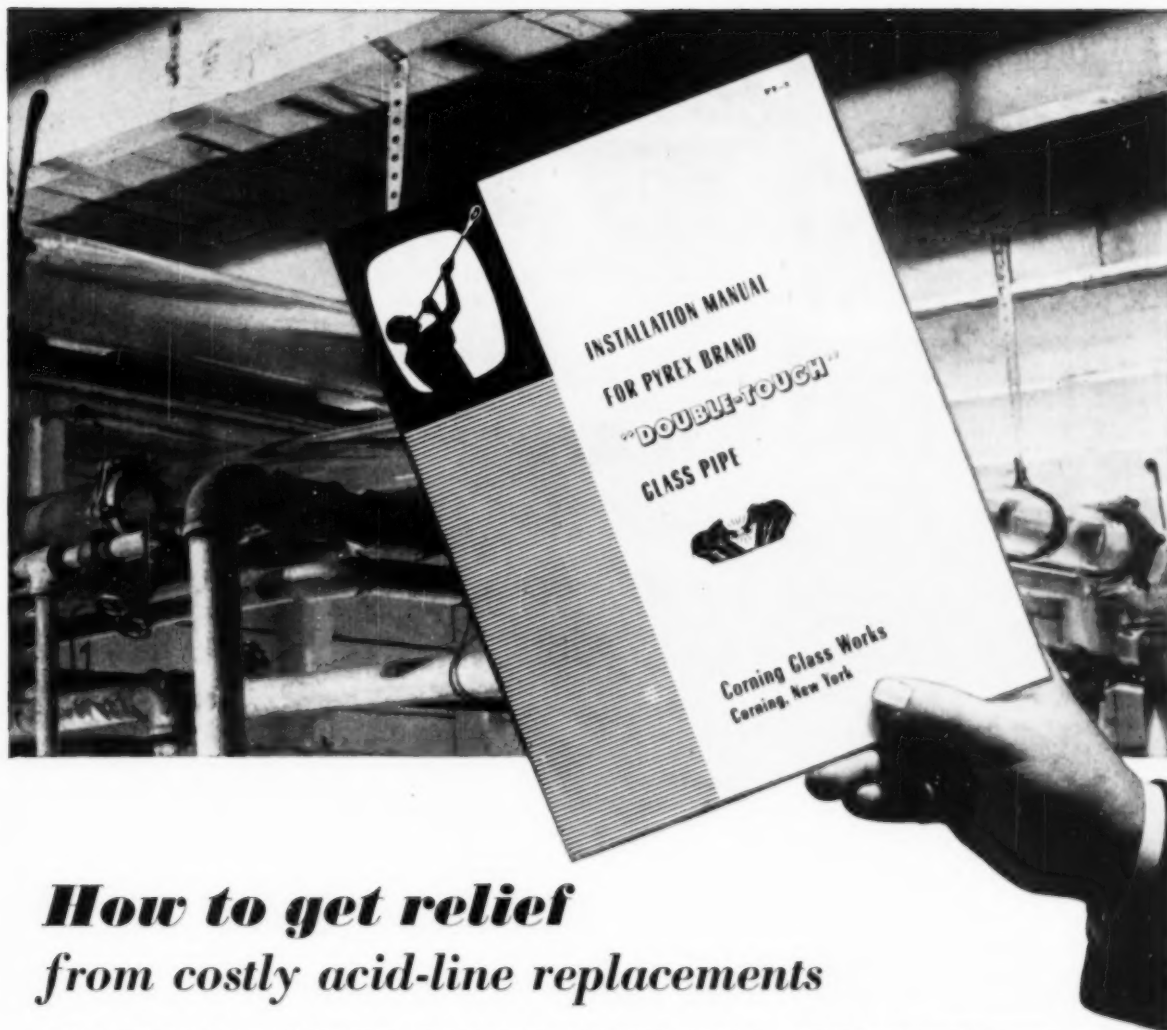
How to get relief
from costly acid-line replacements

This Installation Manual shows how to get full value out of the unequalled corrosion-resistant and non-contaminating properties of PYREX brand "Double-Tough" glass pipe.

The simple installation procedures it describes have been forged from the experience of many chemical and food processing companies who have found relief from excessive acid-line replacement costs in PYREX pipe.

This manual shows how to design and hang PYREX lines in a practical way. It takes you through every step of the various installation practices.

It will answer the questions you are most likely to have in mind about the everyday utility and safety of PYREX pipe. The coupon is for your convenience. We'll be glad to send you a copy free.

What this manual tells you

This Installation Manual tells what you need to know about the thermal expansion, operating pressures and temperatures, and flexibility of PYREX "Double-Tough" glass pipe.

It describes in detail the recom-

mended methods of supporting PYREX pipe and assembling joints.

It illustrates the simple methods to take care of pressure surges in the line, vibration, and lines in areas exposed to traffic.

It details plumbing techniques any mechanic can quickly master, including connecting PYREX pipe to other materials.

Some simple procedure described in this manual may show you the way out of expensive maintenance problems. Send for a copy today.

CORNING GLASS WORKS
CORNING, NEW YORK

Corning means research in Glass

MARGINAL NOTES
(Continued from page 48)

As Others See Us

Chemical Apparatus in the U.S.A. Report by a group of European experts, published by Organization for European Economic Cooperation, Paris (June, 1952), 224 pp. $6.00.

Reviewed by W. R. Marshall, Jr., Associate Dean, College of Engineering, University of Wisconsin, Madison, Wis.

In 1950 the Council of the Organization for Economic Cooperation (O.E.E.C.) dispatched to the United States a Technical Assistance Mission to survey the methods employed in the design and construction of chemical installations and to make a sample study of the development and utilization of control instruments, of the special maintenance methods employed in the chemical industry, and of the resistance of various construction materials to chemical agents. After a little more than a month in the United States and visits to twenty companies, the Mission presented a two-part report, the first section dealing with chemical engineering, the second with well-known processing developments.

The picture of chemical engineering in the United States as seen through the eyes of European engineers is of particular interest. It includes chemical engineering as taught in this country, the organization and importance of chemical engineering construction firms, standardization in chemical engineering, and a general consideration of working methods. Part I concludes:

The complex machinery of the United States chemical industry can operate efficiently only if it is manned by well-trained technicians able to carry out the many specialist functions and provide the necessary co-ordination. The most important is the chemical engineer, whose training was dealt with in some detail in a previous chapter. It will suffice here to recall that university training is broad and quantitative rather than descriptive. Specialist training is provided largely by industry.

Part II, which treats in a descriptive manner processing developments such as the fluidized catalyst process, hypersorption, extractive distillation, and the latest heat-transfer techniques and considers the status of process instrumentation and of education in instrumentation, holds less interest for engineers in the United States although the comparisons of European practices and use of equipment in industry with American are interesting. The reviewer does not believe, however, that these comparisons and the material in Part I are sufficient to make the book a "must" for the library of the chemical engineer.

Here's the low cost way to get chilled water

(FOR EITHER COMFORT OR PROCESS)
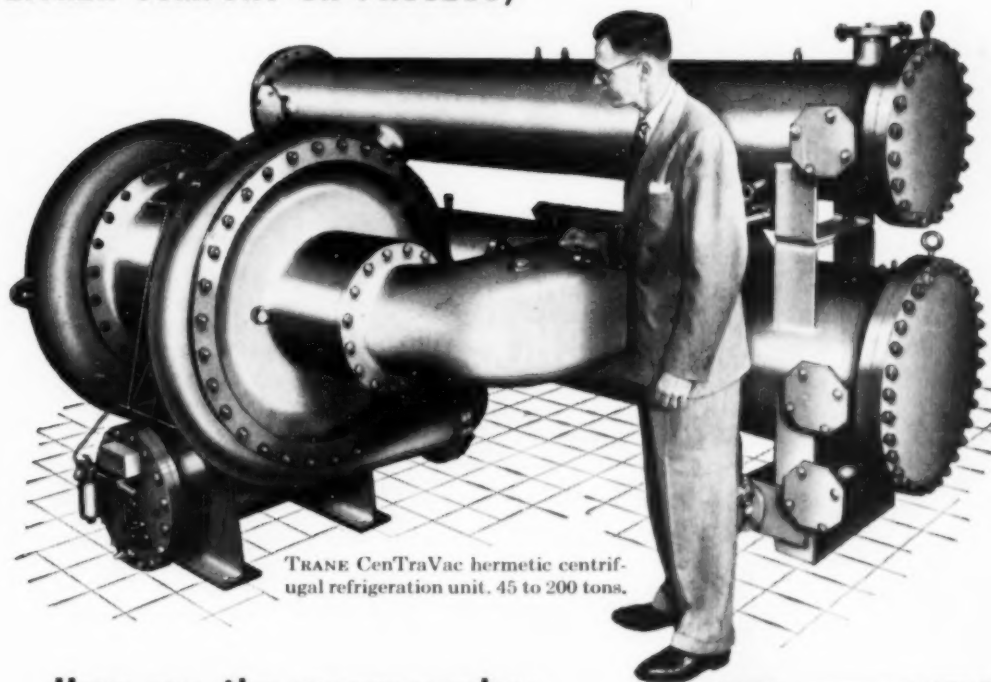
TRANE CenTraVac hermetic centrifugal refrigeration unit. 45 to 200 tons.

Here are the reasons why:

1. **Brings big-job economies to all install-ations** . . . for the first time, the proved money-saving advantages of centrifugal compressors are available to small and medium-sized jobs . . . down to 45 tons.

2. **Complete in one package** . . . self-contained . . . all components are perfectly integrated, therefore operate more efficiently (at less cost).

3. **Easier to install** . . . light, compact . . . needs no special base, only simple water and electrical connections. Saves on installation costs.

4. **Completely automatic** . . . needs no special attention . . . a push-button operation. Frees attendant for other duties.

5. **Fewer moving parts** mean fewer breakdowns, less maintenance. No shaft seals or gear boxes at all. Only *one* major moving part.

6. **Built-in automatic capacity control** permits operation down to 10% of rated capacity . . . saves power.
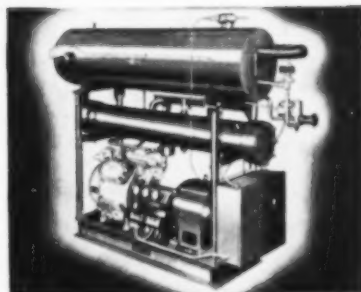
Here's the answer for 10 to 50-ton jobs! A completely packaged water chiller . . . ready to install. The new TRANE Cold Generator comes completely assembled, wired, piped, refrigerant-charged.

TRANE CenTraVac cuts chilled water costs 6 ways

NEWS

DISCUSSION AT CASE ON CORROSION

During the week of Aug. 24 the Department of Chemistry and Chemical Engineering, Case Institute of Technology, Cleveland, Ohio, will conduct a series of lectures and discussions related to corrosion and prevention of corrosion metals in hot water. Included among the subjects to be presented are: electrochemistry of corrosion, influence of temperature on corrosion rates, permeability of plastic coatings (films) to hot water, fundamentals of heat transfer.

The registration fee is $50 and application for attendance may be made by writing to Chairman, Department of Chemistry and Chemical Engineering, Case Institute of Technology, Cleveland 6, Ohio.

PLATE EFFICIENCIES RESEARCH CONTINUING

The research project on plate efficiencies in fractionating towers, cooperatively sponsored by the A.I.Ch.E. and twenty-five chemical, petroleum, and engineering companies, has successfully completed its first active year. This research project, on a 5-year plan, will cover a basic investigation of the fundamentals of tray efficiency both in fractionation and absorption. For the scholastic year 1953-1954, the participating schools will be University of Delaware, University of Michigan and North Carolina State College.

More details of the program, along with the names of the companies contributing to the program and the names of the committee members, were given in "C.E.P.," August, 1952, page 30.

OFFICERS OF CHEMICAL INSTITUTE OF CANADA

Officers of the Chemical Institute of Canada for 1953-1954 elected at the Institute's annual conference in Toronto are:

President—J. W. T. Spinks, F.C.I.C., Chemistry Department, University of Saskatchewan, Saskatoon, Sask.

Vice-President—E. R. Rowzee, F.C.I.C., Polymer Corp., Limited, Sarnia, Ont.

Chairman of the Board—Ludvig Firing, M.C.I.C., Canada Titanium Pigments, Ltd., Montreal, Que.

Treasurer—T. H. G. Michael, M.C.I.C., Director of Research & Development, Howards & Sons (Canada), Ltd., Cornwall, Ont.

General Manager and Secretary—Garnet T. Page, M.C.I.C., Ottawa, Ont.

The 1954 annual conference will be held in Toronto, Ont., on June 21-24.

NEWS

NEW DEVELOPMENTS IN TITANIUM

The first industrial laboratory in the United States to be devoted exclusively to research and development on titanium was recently opened in Niles, Ohio, by Mallory-Sharon Titanium Corp. The new laboratory will be concerned with basic research, development of titanium and titanium alloys, and production testing.

Mallory-Sharon also has announced development of a new melting process for the production of titanium ingots, which increases the yield of metallic titanium from its sponge raw material, now in critical short supply.

Previously two different methods have been used in titanium production: arc melting and induction melting. Arc-melted ingots are not sufficiently homogeneous, and induction-melted ingots have an undesirable carbon content. Mallory-Sharon claims for its new S method that (1) ingots are homogeneous: there are no variations in a sheet, for example, rolled from a single ingot; (2) carbon content can be controlled and held as low as the amount inherent in the sponge material, an electrode of titanium instead of carbon being used and the melting accomplished in a water-cooled copper crucible; (3) more metallic titanium can be produced from a given amount of sponge raw material, since scrap loss is reduced.

FIELD TRIP FOR CORROSION ENGINEERS

A corrosion tour of oil-field equipment will be held from Sept. 30 to Oct. 2, 1953, by the National Association of Corrosion Engineers, according to Tom Matthews, of the publicity committee. Tanks and pumps will be opened up, pipes and sucker rods exposed, and a complete tour of Cosden Refinery, Big Spring, Texas, will be conducted. Further information may be obtained from John A. Knox, The Western Co., Box 310, Midland, Texas.

SYNTHETIC-RUBBER BUY DEPENDS ON BUTADIENE

The projected sale of government-owned synthetic rubber plants to private industry will give rise to complex economic problems, according to T. Ellwood Webster, president, Catalytic Construction Co. "Major among these problems is the subject of butadiene," he said, "since profitable operation of rubber (copolymerization) plants will depend, under private ownership, on an assured supply of low-priced butadiene."

At present according to an industry-wide survey conducted by Catalytic Construction Co. most of the butadiene plants are owned by the Government, all but two using petroleum feed stocks, which are about three times less expensive than alcohol feed stocks. The two alcohol butadiene plants have been periodically operated when butadiene output from the petroleum plants fell below demand and are currently in operation on a reduced-output basis. The Government's synthetic-rubber price has been based on average production costs, and thus the greater expense of alcohol-plant yields has been mitigated by the wider use of petroleum butadiene.

Annual capacity of Government plants is 539,200 short tons of butadiene from petroleum feed stocks and 215,000 tons from alcohol. There seems to be general agreement, the survey notes, that synthetic-rubber requirements, now slightly under 800,000 long tons annually, will increase.

The need for additional butadiene plants, according to the survey, is complicated by two factors: (1) it is not known by whom and for what price the Government's butadiene plants will be purchased, and (2) petroleum butadiene may be produced from either of two by-products of petroleum refining, butylene and butane. Six of the present plants produce butadiene from butylene, and two use butane. Because butylene is used for aviation gasoline, it may not, depending on world conditions, be readily available for privately owned butadiene plants. Butane, however, has in the past been available in abundance and at lower prices. The six plants producing butadiene from butylene, therefore, would require modification if the use of butane as a feed stock became desirable under private ownership.

ATOM NUCLEUS DESCRIBED AS FUZZY

A new description of the nucleus of an atom was provided recently by two physicists at Columbia University. Experiments by James Rainwater and Val Fitch indicate that the nucleus is dense at the center and diffuse toward the edges; in other words, fuzzy. Heretofore scientists have thought of the atom nucleus as a sphere of uniform density with sharply defined edges.

Measurements by Rainwater and Fitch using μ-mesons produced by Columbia's 385,000,000-electron-volt synchrocyclotron indicated a nuclear radius 15% smaller than that found in earlier experiments, and Professor Rainwater believes that the discrepancy is explained by assuming the true structure of the nucleus to be fuzzy edged. Sizing a nucleus may therefore be difficult.

FUTURE MEETINGS AND SYMPOSIA OF A.I.Ch.E.

Chairman of the A.I.Ch.E. Program Committee

Loren P. Scoville, Jefferson Chemical Co.   260 Madison Ave., New York 16, N. Y.

MEETINGS

Annual—St. Louis, Mo., Hotel Jefferson, Dec. 13-16, 1953.

TECHNICAL PROGRAM CHAIRMAN: J. J. Healy, Jr., Assist. to V. P., Monsanto Chemical Co., 1700 S. 2nd St., St. Louis 4, Mo.

Washington, D. C., Statler Hotel, March 8-10, 1954.

TECHNICAL PROGRAM CHAIRMAN: George Armistead, Jr., Consult. Chem. Eng., George Armistead & Co., 1200 18th St. N.W., Washington 6, D. C.

Springfield, Mass., Hotel Kimball, May 16-19, 1954.

TECHNICAL PROGRAM CHAIRMAN: E. B. Fitch, Asst. to Res. Dir., The Dorr Co., Westport, Conn.

Ann Arbor, Mich., Univ. of Mich., Ann Arbor, Mich., June 20-25, 1954—Conference on Nuclear Engineering.

TECHNICAL PROGRAM CHAIRMAN: D. L. Katz, chairman, Dept. of Chem. and Met. Eng., Univ. of Mich., 2028 E. Eng. Bldg., Ann Arbor, Mich.

Glenwood Springs, Colo., Hotel Colorado, Sept. 12-16, 1954.

TECHNICAL PROGRAM CHAIRMAN: Prof. C. H.

Prien, Dept. of Chem. Eng., Univ. of Denver, Denver 10, Colo.

Annual—New York, N. Y., Statler Hotel, Dec. 12-15, 1954.

TECHNICAL PROGRAM CHAIRMAN: G. T. Skaperdas, Assoc. Dir., Chem. Eng. Dept., M. W. Kellogg Co., 225 5Broadway, N. Y. 7, N. Y.

ASST. CHAIRMAN: N. Morash, Titanium Div., National Lead Co., P. O. Box 58, South Amboy, N. J.

Louisville, Ky., March 20-23, 1955.

TECHNICAL PROGRAM CHAIRMAN: R. M. Reed, Tech. Dir., Gas Proc. Div., The Girdler Corp., Louisville 1, Ky.

Houston, Texas, May, 1955.

TECHNICAL PROGRAM CHAIRMAN: J. L. Franklin, Res. Assoc., Humble Oil & Refining Co., P. O. Box 1111, Bayytown, Texas.

Lake Placid, N. Y., Lake Placid Club, Sept. 25-28, 1955.

TECHNICAL PROGRAM CHAIRMAN: L. J. Coulthurst, Chief Proc. Designer, Foster Wheeler Corp., 165 Broadway, New York 6, N. Y.

Detroit, Mich.—Statler Hotel, Nov. 27-30, 1955.

SYMPOSIA

Distillation

CHAIRMAN: D. E. Holcomb, Dean of Eng., Texas Technological College, Lubbock, Tex.
MEETING—St. Louis, Mo.

Dust and Mist Collection

CHAIRMAN: C. E. Lapple, Dept. of Chem. Eng., Ohio State University, Columbus 10, Ohio.
MEETING—St. Louis, Mo.

Drying

CHAIRMAN: L. E. Stout, Dept. of Chem. Eng., Washington Univ., St. Louis 5, Mo.
MEETING—St. Louis, Mo.

Use of Computers in Chemical Engineering

CHAIRMAN: John R. Bowman, Head, Dept. of Res. in Phys. Chem., Mellon Institute of Industrial Research, Pittsburgh 13, Pa.
MEETING—St. Louis, Mo.

Heat Transfer

CHAIRMAN: D. L. Katz, Chairman (Address: See Ann Arbor Meeting).
MEETING—St. Louis, Mo.

Carbonization

CHAIRMAN: R. S. Rhodes, Asst. Mgr., Prod. Dept., Koppers Co., Inc., Koppers Bldg., Pittsburgh 19, Pa.
MEETING—St. Louis, Mo.

Industrial Waste Disposal

CHAIRMAN: K. S. Watson, Coordinator, Waste Treatment, Mfg. Facilities Serv. Dept., Gen'l. Electric Co., 202 State St., Schenectady, N. Y.
MEETING—St. Louis, Mo.

Mixing

CHAIRMAN: J. H. Rushton, Dept. of Chem. Eng., Illinois Inst. of Tech., Chicago, Ill.
MEETING—Washington, D. C.

Patents

CHAIRMAN: W. C. Asbury, V. P., Std. Oil Dev. Co., 15 W. 51st St., New York 19, N. Y.
MEETING—Washington, D. C.

Chemical Engineering in the Fertilizer Industry

CHAIRMAN: G. L. Bridger, Head, Dept. Chem.

& Mining Eng., Iowa State College, Ames, Iowa.
MEETING—Washington, D. C.

Liquid Entrainment and Its Control

CHAIRMAN: H. E. O'Connell, Ethyl Corp., P. O. Box 341, Baton Rouge, La.
MEETING—Washington, D. C.

Nuclear Engineering

CHAIRMAN: D. L. Katz, Chairman (Address: See Ann Arbor Meeting).
MEETING—Ann Arbor, Mich.

Reaction Kinetics

CHAIRMAN: N. R. Amundson, Dept. of Chem. Eng., Univ. of Minnesota, Minneapolis 14, Minn.
MEETING—New York, N. Y.

Gas Absorption

CHAIRMAN: R. L. Pigford, Div. of Chem. Eng., Univ. of Delaware, Newark, Del.
MEETING—New York, N. Y.

Centrifugation

CHAIRMAN: J. O. Maloney, Chairman, Dept. Chem. Eng., Univ. of Kansas, Lawrence, Kan.

Nucleation Processes

CHAIRMAN: E. L. Piret, Dept. Chem. Eng., Univ. of Minn., Minneapolis 14, Minn.

Agglomeration

CHAIRMAN: A. P. Weber, International Engineering, Inc., 15 Park Row, New York, N. Y.

Submitting Papers

Members and nonmembers of the A.I.Ch.E. who wish to present papers at scheduled meetings of the Institute should follow the following procedure. First, write to the Chairman of the A.I.Ch.E. Program Committee, address at the top of this page. A carbon of the letter sent to the Chairman should also go to the Technical Program Chairman of the meeting at which you wish to present the paper. Still another carbon should go to the Editor of Chemical Engineering Progress, F. J. Van Antwerpen, 120 East 41st St., New York 17, N. Y.

If the paper is scheduled for a symposia the procedure is somewhat different. A carbon of the letter should go to the Chairman of the Symposia instead of the Chairman of the Technical Program, since symposia are scheduled for meetings only when they are complete and approved by the Chairman of the National Program Committee.

Before Writing the Paper

Before beginning to write your paper you should obtain from the meeting Chairman, or from the office of the Secretary of the A.I.Ch.E., at 120 East 41st St., New York, a copy of the A.I.Ch.E. Guide to Authors, and Guide to Speakers. These cover the essentials required for submission of papers to the A.I. Ch.E. or its magazines.

Copies of Manuscript

Five copies of each manuscript must be prepared. For meetings, one should be sent to the Chairman of the symposium, and one to the Technical Program Chairman of the meeting at which the symposium is scheduled. If no symposium is involved, the two copies should be sent to the Technical Program Chairman. The other copies should be sent to the Editor's office since manuscripts are automatically considered for publication in Chemical Engineering Progress, or the symposium series of Chemical Engineering Progress, but presentation at a meeting is no guarantee that they will be accepted.

Deadlines for Meetings

DEADLINE DATES

ST. LOUIS MEETING—August 13, 1953

WASHINGTON, D. C. MEETING—November 8, 1953

SPRINGFIELD MEETING—January 9, 1954

ANN ARBOR MEETING—February 15, 1954

GLENWOOD SPRINGS MEETING—May 12, 1954

NEW YORK MEETING—August 12, 1954

LOUISVILLE MEETING—November 20, 1954

HOUSTON MEETING—definite dates have not been set.

LAKE PLACID MEETING—definite dates have not been set.

DETROIT MEETING—July 27, 1955

Secretary's Report

S. L. TYLER

As it was impractical to hold a formal meeting of the Executive Committee in July, the necessary actions were handled by mail ballot. The Minutes of the Executive Committee Meeting of June 15 were approved. Candidates whose names appeared in the June issue of "C.E.P.," with one exception, were elected to the grades of membership indicated. In addition to this list, twenty-nine Student members were elected, which brings the total Student membership to 1,808.

R. P. Milford was appointed representative of the Knoxville-Oak Ridge Section replacing J. Shacter on the Membership committee; A. P. Weber was appointed a member of the Program Committee.

Two elections were rescinded because of nonacceptance.

Two memberships were discontinued because of nonpayment of dues, and three resignations were accepted.

J. S. McClure and W. T. Woodmencey were placed on the Suspense List because of having entered the Armed Forces.

E. A. Brown, L. A. Chabot, J. J. Foster, B. F. Schrader, Jr., and D. E. Wollscheid have been removed from the Suspense List since they had completed their tour of duty in the Armed Forces.

CANDIDATES

(*Continued from page* 34)

Mergens, Edward Harold, Queens Village, N. Y.
Meyer, Richard D., Louisville, Ky.
Michel, Philip R., Saginaw, Mich.
Milholland, Dick E., Fort Wayne, Ind.
Miller, Edward, Bronx, N. Y.
Mingle, John O., Oakley, Kan.
Minushkin, Bertram, Brooklyn, N. Y.
Montanye, Harry E., Philadelphia, Pa.
Mosier, Chas. F., Jr., Middletown, Ohio
Myers, Hal H., Shaker Heights, Ohio
Neesbye-Hansen, O., New Castle, Del.
Niehaus, Robert F., Cincinnati, Ohio
Nodorp, Richard W., Arlington Heights, Ill.
O'Brien, Donald L., So. St. Paul, Minn.
Oldham, Arthur F., LeRoy, N. Y.
Oschell, Robert G., Philadelphia, Pa.
Paddock, Charles, Butler, N. J.
Paredes, Edward A., Highland Park, N. J.

Parente, Emil J., Long Island City, N. Y.
Parkes, John D., Niagara Falls, N. Y.
Pennepacker, Jesse G., Jr., New Hyde Park, N. Y.
Peoples, Leo Clark, Columbus, Ohio
Prendergast, Harry W., Manchester, N. H.
Price, Jerry L., LaMarque, Tex.
Pugh, Ralph A., Richland, Wash.
Rankin, Sidney, Baltimore, Md.
Rathmann, Clyde L., Corning, N. Y.
Reed, John Charles, Newark, Del.
Reinhart, Richard D., Austin, Minn.
Reinmuller, Ernest, Dayton, Ohio
Riley, Frank G., Spring Gap, Md.
Rinker, John R., Terre Haute, Ind.
Risch, Robert A., Milwaukee, Wisc.
Ritter, Robert L., Louisville, Ky.
Rodowsky, John Charles, Baltimore, Md.
Rome, Richard M., Rockland, Mass.
Rose, Vernon A., Audubon, Iowa
Rosenbluth, Murray J., Philadelphia, Pa.
Rue, Loring W., Jr., Paulsboro, N. J.
Rule, Harold E., Kingsport, Tenn.
Ryan, Francis J., Philadelphia, Pa.
Saile, George O., Cincinnati, Ohio
Sandler, Henry J., Philadelphia, Pa.
Sashihara, Thomas F., E. Cleveland, Ohio
Savage, Howard H., McAlester, Okla.
Schlarb, James A., Brazil, Ind.
Sharp, Hugh T., New York, N. Y.
Shulko, Richard M., Binghamton, N. Y.
Simms, David M., Taunton, Mass.
Simpson, Hugh C., Cambridge, Mass.
Sjogren, Roy G., Philadelphia, Pa.
Skorko, Joseph J., Camden, N. J.
Skrivan, Joseph F., Baltimore, Md.
Smith, Theodore R., Charleston, W. Va.
Snyder, Marshall T., Drexel Hill, Pa.
Spillane, Robert G., Hastings, Mich.
Squires, Paul H., Madison, Wisc.
Stahl, Albert, Cincinnati, Ohio
Stein, Seymour Samuel, Richmond, Va.
Stewart, William J., Indianapolis, Ind.
Stone, Francis X., Gibbstown, N. J.
Subtelney, Daniel G., Springfield, Mass.
Swanson, W. M., LaGrange, Ill.
Tartaron, Garland J., Ishpeming, Mich.
Thomas, Richard I., Webster Groves, Mo.
Townsend, Robert W., Brockton, Mass.
Truscott, Thomas B., Huntsville, Ala.
Unger, Harold, Bronx, N. Y.
Vershon, Seymour M., Jr., Holyoke, Mass.
Vogel, Harry, Jamaica, N. Y.
Wohlmann, Gilbert E., St. Louis, Mo.
Waters, C. William, Clarksburg, W. Va.
Weaver, John W., Haddonfield, N. J.
Weaver, Robert Edgar, New Orleans, La.
Weight, George W., Evanston, Ill.
White, Lionel N., Nashua, N. H.
Whitsel, Calvin V., Borger, Tex.
Wielinski, Lawrence A., Swanton, Ohio
Wiesenborn, Robert Fred, St. Louis, Mo.
Wilczynski, Philip T., Nekoosa, Wisc.
Williams, Glen A., LaPorte, Ind.
Wilson, James E., Jr., Belfast, Maine
Wilson, Stafford L., Fort Sam Houston, Tex.
Wittkopf, Warren A., Seattle, Wash.
Wong, Franklin K., Clarkston, Wash.
Wood, James E., Jr., West Chester, Pa.
Wszolek, Edward J., Springfield, Mass.
Zbinden, Harry W., Cranford, N. J.
Zwicky, John F., Richmond Hill, N. Y.

VISCOSITY-DENSITY DATA ISSUED BY A.S.M.E.

Comprehensive data on the viscosity and density of more than forty lubricating fluids at temperatures from 32° to 425° F. and at pressures up to 150,000 lb./sq.in. have been collected in two volumes entitled "Pressure-Viscosity Report" issued by the A.S.M.E.

The pressure-temperature-viscosity-density measurements were carried out at Harvard University under the advice of a board composed of experts from universities, government laboratories, and the petroleum, synthetic lubricant, and equipment-building industries. These measurements have been entered on IBM punch cards, sets of which may be obtained from the A.S.M.E. at a nominal charge.

The complete data are also available from the A.S.M.E., 29 West 39 Street, New York 18, N. Y., at a cost of $10 for the two volumes.

EMPLOYEES' FUND BUYS CONTROL OF A. D. LITTLE

Controlling interest in Arthur D. Little, Inc., was acquired recently by the employees' retirement trust, Earl P. Stevenson, president, announced. The company, Mr. Stevenson stated, is "composed largely of professional personnel. Since our success as an organization depends in part upon the creativity of our staff, it is fitting that the staff should have a stake in the Company's corporate success."

Among the shares of common stock bought by the employees' fund were those held in trust for the benefit of the Massachusetts Institute of Technology. Proceeds from the sale, which amounted to $1,300,000, will be retained in the existing Arthur Dehon Little Fund for the benefit of M.I.T., and it is expected, Institute officials said, that support of educational and research projects chiefly in the departments of chemical engineering and chemistry will be continued.

CHEMICAL ENGINEERING FEDERATION FOR EUROPE

A European Federation of Chemical Engineering was inaugurated in Paris on June 20, 1953, for the purpose of promoting European cooperation in chemical engineering and equipment. Scientific and technical societies from Spain, Yugoslavia, Germany, Finland, The Netherlands, Portugal, France, and Switzerland were represented at the inauguration. Headquarters of the federation are at Maison de la Chimie, 28 Rue Saint-Dominique, Paris, and Dechema-Haus, Frankfort am Main, Rheingauallee, 25.

Wednesday, Sept. 16, 1953

TECHNICAL SESSION NO. 9

Symposium on Ion Exchange

9:00 A.M.—FUNDAMENTAL FACTORS IN ION EXCHANGE PROCESS DESIGN, W. C. Bauman, Dow Chemical Co., Midland, Mich.

9:40 A.M.—MIXED BED DEIONIZATION AT HIGH FLOW RATES, J. R. Caddell and R. L. Moison, Du Pont Co., Wilmington, Del.

10:10 A.M. — A PRACTICAL SEPARATION YTTRIUM GROUP RARE EARTHS FROM GADO-LINITE BY ION EXCHANGE, F. H. Spedding and J. E. Powell, Iowa State College, Ames.

10:40 A.M.—CONTINUOUS COUNTERCURRENT ION ADSORPTION AND STRIPPING USING ION-EXCHANGE RESINS, W. W. Koenig, A. L. Babb, and Joseph L. McCarthy, University of Washington, Seattle, Wash.

11:10 A.M.—GENERAL NUMERICAL SOLUTION FOR SOLID DIFFUSION IN FIXED BEDS, J. B. Rosen, Forrestal Research Center, Princeton University, Princeton.

TECHNICAL SESSION NO. 10

General Technical Program

9:30 A.M.—HEAT TRANSFER AND FLOW FRIC-TION CHARACTERISTICS OF POROUS MEDIA, J. E. Coppage and A. L. London, Stanford University, Stanford, Calif.

10:00 A.M.—HEAT TRANSFER TO GRANULAR MATERIALS—SETTLED BEDS MOVING DOWN-WARD OVER HORIZONTAL TUBES, J. T. Moody, R. L. Pigford, and N. A. Copeland, Humble Oil & Refining Co., Baytown, Tex.

10:30 A.M.—NATURAL CONVECTION HEAT TRANSFER AT REDUCED PRESSURES, J. R. Kyte, A. J. Madden, and Edgar L. Piret, University of Minnesota, Mineapolis.

11:00 A.M.—FREE CONVECTION MASS TRANS-FER AT VERTICAL PLATES, C. R. Wilke, C. W. Tobias, and Morris Eisenberg, University of California, Berkeley.

TECHNICAL SESSION NO. 11

Symposium on Ion Exchange

2:00 P.M.—ENGINEERING EVALUATION OF MOVING AND FIXED BED ION EXCHANGE PROCESSES, N. K. Heister, R. K. Cohen, and R. C. Phillips, Stanford Research Institute.

2:30 P.M.—HYDROXIDE CYCLE OPERATIONS WITH STRONGLY BASIC ANION EXCHANGE RESINS, R. M. Wheaton, Dow Chemical Co.

3:00 P.M.—COUNTERCURRENT ION EXCHANGE WITH GROSS COMPONENTS, I—IN AN EQUILIBRIUM STAGE CONTACTOR, N. K. Heister, R. C. Phillips, and R. K. Cohen, Stanford Research Institute, Stanford.

SAN FRANCISCO MEETING

of animals in the adjoining Academy of Sciences Building. A walk through the near-by Japanese Tea Garden and a stop for a cup of tea in the midst of this artistic rock garden, with "moon" bridges arching over quiet lagoons, dwarfed trees, and towering bamboo groves will prove refreshing.

San Francisco's famous beach still holds vestiges of its early history. The Cliff House, historic restaurant perched high above the rocks, is a mecca for newcomers to the city who sit in its glassed-in dining room and watch the sea lions scramble around on the rocks below.

San Francisco is not just a city fashioned for sightseeing and recreation, however. The business section, often called the Wall Street of the West, is a cluster of skyscrapers centered around Montgomery Street. Around the Bay are dotted factories and industrial plants in which the chemical industry is prominently represented.

For many years the San Francisco Bay Area has been an outstanding center of academic institutions. Across the Bay in Berkeley is the University of California. Down the Peninsula is world-famed Stanford University. San Francisco State College is located in the city, as is the University of San Francisco, which was founded in 1855 by the Jesuits. Mills College in East Oakland was founded in 1852 and is the oldest college for women west of the Mississippi.

Plant Trips

The following trips are planned to acquaint the visiting engineers with the chemical industry of the area. They are scheduled for Tuesday, Sept. 15.

1. *Standard Oil Company of California, Richmond plant.*
   University of California, Radiation Laboratory.

The Richmond Refinery of Standard Oil is spread over approximately 1900 acres. More than 1,000 different products are manufactured here from petroleum. A general tour will be made of the refinery including the new crude unit and vacuum flasher, wax plant, packaging plant, and Thermofor catalytic cracker.

A complimentary luncheon will be served in the Standard Oil Company cafeteria.

The Radiation Laboratory of the University of California is one of the best-known centers of atomic research. Visitors will be given an introductory lecture on the various machines and points of interest. The tour will include the cyclotron and the new bevatron.

2. Shell Development Company Laboratories, Emeryville.
   Colgate-Palmolive-Peet Co., Berkeley

Shell's Emeryville research center is the largest organization of its kind on the West Coast. Research and engineering development are carried out on a variety of problems in the fields of petroleum technology and chemical synthesis from petroleum sources. Some eighty of the professional staff can be classified as chemical engineers. The tour will include the pilot plants, instrumentation and glass blowing, fuels and lubricants engineering, plastics application lab.

A complimentary luncheon will be served in the Shell cafeteria.

Visitors to the Colgate-Palmolive-Peet Co. plant will see the processes involved in the manufacture of soap, glycerine, detergents, and toilet articles.

3. Columbia-Geneva Steel division, U. S. Steel Corp., Pittsburg.
   The Dow Chemical Co., Pittsburg.

Occupying 400 acres, the present Pittsburg plant is recognized today as a major producer of steel. The plant tour will include the open-hearth furnaces, rolling mill, rod mill, wire and nail mill, wire rope mill, foundry, and the newly enlarged cold reduction sheet and tin plate mill.

Luncheon will be served at the plant cafeteria.

Visitors to the Western Division plant of The Dow Chemical Co. will be shown facilities in the electrolytic chlorine-caustic soda and caustic potash plants, chlorinated hydrocarbons plant, caustic soda evaporator, liquid chlorine plant, and the ammonia plant.

4. Wineries in the Napa Valley, St. Helena.
   U. S. Department of Agriculture, Western Regional Research Laboratory, Albany.

The Napa Valley produces some of the world's finest dinner wines. The picturesque wineries of Beringer Bros. and Charles Krug will be visited.

A picnic lunch will be served under the old oak trees along with a wine-tasting program.

Visitors to the Western Regional research laboratory will see the engineering laboratories, where new equipment for food processing and other agricultural product processing are developed and tested. Facilities of the food-processing laboratory, where experimental and test samples are packed, will be inspected. Several examples of problems under laboratory development will be shown.

(The End)

LOCAL SECTION NEWS

ANNUAL SYMPOSIUM NEW YORK SECTION

An all-day meeting, a symposium, covering the topics of Management, New Materials of Construction, and Solids Handling will be conducted by the New York Section of the American Institute of Chemical Engineers on Oct. 22, 1953, at the Hotel New Yorker, New York. This Section holds an annual symposium. Last year's one-day meeting which concentrated on Adsorption and Cost Estimation was attended by more than 600 engineers and chemists from the Eastern part of the United States.

Morgan Hoover, Chemical Engineering, is the chairman of the session on Management. Kempton Roll, Lead Industries Association, will conduct on New Materials of Construction, and R. B. Thompson, The Dorr Co., will preside at Solids Handling.

The program of papers now scheduled is as follows:

MANAGEMENT

FINANCING CHEMICAL EXPANSION, W. McElvea, Empire Trust Co.

FUNDAMENTALS OF ORGANIZATION, Ernest Dale, American Management Assoc.

MANAGEMENT CONTROLS, Arthur Tichenor, Booz, Allen and Hamilton Co.

MAINTENANCE, Henry Shockley, Du Pont Co.

MATERIALS HANDLING, A. Strong, American Cyanamid Co.

UTILIZATION OF PERSONNEL, John Burns, American Management Assoc.

NEW MATERIALS OF CONSTRUCTION

THE TREND TO THICK PROTECTIVE COATINGS FOR CHEMICAL EXPOSURES, R. S. Pierce, Penn Salt Manufacturing Co.

NEW DEVELOPMENT IN PLASTICS FOR CHEMICAL CONSTRUCTION, G. S. Laaft, The Bolta Co.

THE APPLICATION OF TITANIUM AND ZIRCONIUM METALS AND THEIR ALLOYS IN CHEMICAL CONSTRUCTION, G. C. Kiefer, Allegheny Ludlam Steel Corp.

SOLIDS HANDLING

CONVEYING — A GENERAL SURVEY, Harold Weil, Chain Belt Co.

NEW DEVELOPMENTS IN CONVEYING, R. V. Jackson, Hewitt Robins Co.

SOLIDS HANDLING AT NICARO, CUBA (Speaker to be announced)

ALL-DAY MEETING IN PULLMAN, WASH.

The Columbia Valley Section was host at a recent all-day meeting at the New Student Union Hall, Pullman, Wash. The order of exercises for the day included a banquet at which A. A. Janos, General Electric Co., Hanford Atomic Products Operation, Richland, Wash., was master of ceremonies. The speaker at the banquet was George T. Austin, head, chemical engineering department, Washington State College. His talk was titled "Modern Chemical Engineering Education."

The papers presented were as follows:

A STUDY OF THE VARIABLES DETERMINING THE FLOWRATE OF GRANULAR MATERIALS THROUGH A HORIZONTAL ORIFICE, F. C. Franklin and L. N. Johanson, department of chemical engineering,, University of Washington, Seattle, Wash.

AN APPROACH TO CYCLONE DESIGN, W. S. Munro, Monsanto Chemical Co., Seattle.

SOME ASPECTS OF CHEMICAL ENGINEERING IN NUCLEAR PLANTS, W. J. Dowls (Presented by W. D. Ambrose), General Electric Co., Hanford Atomic Products Operation, Richland.

UTILIZATION OF SYNTHETIC LUBRICANTS IN AIRCRAFT, F. G. Dean,, Boeing Airplane Co., Seattle.

TIME, MOTION, AND METHOD CONSIDERATIONS FOR CHEMICAL PLANTS, K. A. Clark, General Electric Co., Hanford Atomic Products Operation, Richland.

CALCULATION OF CORROSION RATES FROM CATHODIC PROTECTION CURRENTS, P. A. Johnson and A. L. Babb, department of chemical engineering, University of Washington, Seattle.

TEFLON WELDING, H. G. Henry, General Electric Co., Hanford Atomic Products Operation.

INERT GAS SHIELDED TUNGSTEN ARC WELDING, E. B. LaVelle, General Electric Co., Hanford Atomic Products Operation, Richland.

NEW GROUP TO FORM IN NAUGATUCK VALLEY

A group of chemical engineers is now in the process of forming a chemical engineers' club in the New Haven and Naugatuck Valley area with the intent of affiliating eventually as a local section with the A.I.Ch.E. An organizational meeting will be held in September, 1953, at the Sterling Chemistry Laboratory of Yale University.

Those who may be interested can contact the following persons for further information: Shepherd Lippa, Sponge Rubber Products Co., Shelton, Conn.;

R. W. Southworth, department of chemical engineering, Yale University; Alan T. Lincoln, American Cyanamid Co., Wallingford, Conn.; and Walter S. Kaghan, and Joseph J. Levitzky, Olin Industries, Inc., New Haven.

The final dinner meeting of the year of the Twin Cities Section, a party especially for the wives, was held at Napoleon's in St. Paul, Minn. Corsages for the ladies were furnished by Minnesota Mining & Manufacturing Co., and other gifts were supplied by that company and Rayette Inc., Economics Laboratory, and Pillsbury Mills, Inc. Ninety-eight engineers and wives attended the dinner and listened to Helen Hallbert, home economist from General Mills' Betty Crocker Kitchen, discuss one of the most important kitchen tools, baking pans. It was scientific enough to interest the engineers and practical enough to interest the wives, according to the reporter, R. W. Batey.

Forty-five members, Central Virginia Section, their guests, and their wives enjoyed an outing and dinner held at the Keswick Country Club, Charlottesville, Va. Professor John A. Rorer, department of education, University of Virginia, talked of his visit to England and British West Africa. J. H. Mallinson, reporting the meeting, said that the audience especially enjoyed the picturesque details of educational institutions as they exist in British West Africa.

TECHNICAL PROGRAM
(Continued from page 61)

3:30 P.M.—ION EXCHANGE EQUILIBRIA OF SODIUM CHLORIDE, POTASSIUM-CHLORIDE, WATER, AND DOWEX 50, J. A. Whitcombe, J. T. Banchero, and R. R. White, University of Michigan, Ann Arbor.

4:00 P.M.—EXPERIMENTAL INVESTIGATIONS OF ION EXCHANGE MECHANISMS IN FIXED BEDS BY MEANS OF AN ASYMPTOTIC SOLUTION, L. Lapidus and J. B. Rosen, Forrestal Research Center, Princeton.

TECHNICAL SESSION NO. 12
General Technical Program

2:00 P.M.—LIQUID FILMS IN VISCOUS FLOW, Melbourne L. Jackson, University of Colorado, Boulder, Colo.

2:45 P.M.—FACTORS AFFECTING GAS RECIRCULATION AND PARTICLE EXPANSION IN SPRAY DRYING, J. A. Buckham and R. W. Moulton, University of Washington, Seattle, Wash.

3:15 P.M.—ABSORPTION OF HYDROCHLORIC ACID IN A WETTED-WALL ABSORBER, C. J. Dobratz, R. J. Moore, R. D. Barnard, and R. H. Meyer, Dow Chemical Co., Pittsburg, Calif.

3:45 P.M.—A GENERAL GRAPHICAL METHOD FOR FOUR-COMPONENT EXTRACTION DESIGN, John E. Powers, University of California, Berkeley.

PEOPLE

Anson B. Nixon was elected chairman of the board of Hercules Powder Co. at a recent meeting of the board in Wilmington, Del. Mr. Nixon resigned as vice-president of the company to accept the board chairmanship. Prior to becoming vice-president he was general manager of the naval stores department. He received his B.S. degree from Purdue.

Gustav Egloff, of Universal Oil Products Co., has been appointed a member of the Washington Award Commission of the Western Society of Engineers, 1953-1956; also a national director of the Armed Forces Chemical Association for 1953-1954; and a member of the Advisory Committee, Midwest Chapter, of the Armed Forces Chemical Association, 1953-1954.

Eric R. Braun, formerly director, engineering division, Merck & Co., Inc., has been appointed assistant to the vice-president for operations. In this position he will supervise and coordinate the engineering, purchasing, and traffic activities of the company. Mr. Braun joined the organization in 1937 as a chemical engineer, and since that date has served as manager of the pilot plant, and assistant director of the research and development division in the chemical engineering design department. Subsequently, he was made assistant director of the engineering division and was appointed director in 1950.

John J. Grebe has recently been appointed director of nuclear research and development for The Dow Chemical Co., Midland, Mich. In this capacity Dr. Grebe represents Dow in a joint study with the Detroit Edison Co. and the Atomic Energy Commission looking to the development of a nuclear reactor to produce power.

Dr. Grebe has been a member of the Dow organization since 1924. During World War II he was a consultant to the Office of the Rubber Director. He attended the Oak Ridge National Laboratory's course on applications of nuclear energy in industry. Currently he is a consultant to the Army Chemical Corps.

Gilbert J. Kenngott and **Gerard Butler** have been appointed acrylonitrile department superintendent, and production supervisor respectively at American Cyanamid Co.'s nitrogen chemicals plant near New Orleans, La. Mr. Kenngott, with Cyanamid since 1943, served as a technical assistant in the plastics and resins division's production department prior to his new appointment. He was graduated from Brooklyn Polytechnic Institute with a B.S. degree in chemical engineering. Mr. Butler joined the company in 1951 and was employed by the firm's Lederle laboratories division, Pearl River, N. Y., and later worked at Chemical Construction Corp.

John B. Calkin, chemical engineer and consultant, has been named assistant to the president, Foster D. Snell, Inc., New York. He will also be director of market research. Recently he was associated with the University of Maine where he was director of the department of industrial cooperation and secretary. The University of Maine Pulp & Paper Foundation. He is an associate professor of chemical engineering. Mr. Calkin has been connected with several companies in the pulp and paper industry doing chemical research.

The promotion of **Reuben F. Pfennig** to research specialist has been announced by the research & development division at Baytown refinery, Humble Oil & Refining Co. Mr. Pfennig received his B.S. and M.S. degrees in chemical engineering at the University of Texas. He was first employed in Humble's specialty manufacturing department in Houston (1936), transferring to the research group at Baytown the following year. His work at Baytown has been on various phases of research on distillation, extractive distillation, solvent extraction, etc. In his new position Mr. Pfennig will specialize in the development and evaluation of processes for the manufacture of petrochemicals.

Henry L. Shuldener, president of Water Service Laboratories, Inc., has been elected chairman of the Executive Committee, New York Engineers' Committee on Student Guidance, a group of more than 100 volunteers from the five Founders Societies, i.e., A.I.Ch.E., A.S.C.E., A.S.M.E., A.I.M. & M.E. and A.I.E.E. This group operates under the auspices of the Engineers' Council for Professional Development.
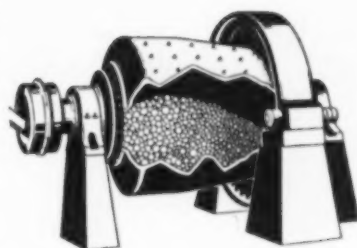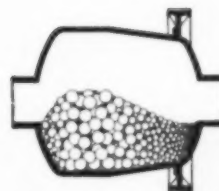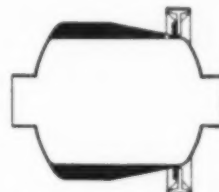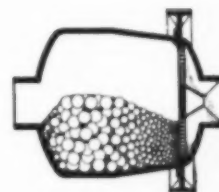
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James K. Rice, vice-president of Cyrus Wm. Rice & Co., Industrial Water Consultants, Pittsburgh, Pa., was among 100 young men honored by Time Magazine and the Pittsburgh Chamber of Commerce. They were chosen on the basis of performance in their chosen field as an indication of their promise as leaders of the future. Mr. Rice received his B.S. and M.S. degrees in chemical engineering from Carnegie Institute of Technology, joining Cyrus Wm. Rice & Co. in 1947.

Louis Gleekman has joined the research and development division of Wyandotte Chemicals Corp., where he will work in the materials engineering section of the chemical engineering research department. At Wyandotte Dr. Gleekman will be concerned with a study of corrosion problems and the evaluation of materials of construction for specific applications in plants and equipment of the company. Dr. Gleekman's past experience includes corrosion and metallurgical research, unit operation studies, and technical service with Sherritt-Gordon Mines, Ltd., Sam Tour and Co., Inc., and Mathiesen Alkali Works, magnesium plant. He has also served as an instructor in metallurgy, unit operations, corrosion, and advanced physical metal metallurgy at the Univ. of Del.

The promotion of **B. Otto Wheeley** to the position of Southern district sales manager, tar products division, Koppers Co., Inc., was recently announced. In his new position he will have charge of sales of all the division's products in ten Southern states. Since joining Koppers in 1943 as a cadet engineer, he has been administrative-assistant, assistant to district manager, and assistant district sales manager.

G. P. Hungerford, research manager, American Machine & Foundry Co., will be transferred to the company's new laboratory to be established in Springdale, Conn., continuing to direct the company's activities in chemical research and development. He joined the company in 1946 as a chemical engineer, becoming senior chemical engineer a few years later. He was formerly associated with Monsanto Chemical Co. and during World War II served with the U. S. Navy.

E. H. Amick, Jr., assistant dean of engineering at Columbia University, and **Francis W. Winn,** process design engineer, have joined Fractionation Research, Inc., to head its technical organization. Dr. Amick assumed his duties as technical director on June 1. Mr. Winn has been employed as associate director.

Dr. Amick received his doctorate in chemical engineering from Yale University. He has served the industry in research and process engineering supervisory capacities, and held chemical engineering teaching and administrative posts at the University of Pennsylvania and Columbia University. He was the United States representative on the committee in Germany working on decartelization of the I. G. Farbenindustrie after World War II.

Winn, a graduate of the University of Oklahoma's chemical engineering school, spent more than four years in the Army Chemical Warfare Service, working during most of this time with the Petroleum Administration for War.

James M. Church was recently promoted to full professor of chemical engineering at Columbia University and will be on sabbatical leave this coming year with the Du Pont Co. He is joining the polychemcials department in their development laboratories at the Experimental Station in Wilmington, Del.

James H. Wiegand has been named to the post of assistant chairman of the the department of chemistry and chemical engineering, Southwest Research Institute, and will play a leading role in the non-profit industrial laboratory's growing program in the fields of high energy materials, including explosives, propellants, pyrotechnics and incendiaries. Dr. Wiegand's experience includes twelve years in propellants and explosives, serving successively with the Du Pont Co., the Army Ordnance Department, the Ballistics Research Laboratory, and the Naval Ordnance Test Station. As associate head of the rockets department at the Naval Station, Dr. Wiegand, who received his doctor of philosophy degree from the University of Michigan, was responsible for all propellant work.

Thompson Chandler, chemical engineer, Carbide & Carbon Chemicals Corp., South Charleston, W. Va., has been nominated for regional vice-president of the American Society of Mechanical Engineers.

(More News about People on page 71)

CLASSIFIED SECTION

Advertisements in the Classified Section of Chemical Engineering Progress are payable in advance, and are placed at 15c a word, with a minimum of four lines accepted. Box number counts as two words. Advertisements average about six words a line. Members of the American Institute of Chemical Engineers in good standing are allowed one six-line insertion (about 36 words) free of charge per year. More than one insertion to members will be made at half rates. In using the Classified Section of Chemical Engineering Progress it is agreed by prospective employers and employees that all communications will be acknowledged, and the service is made available on that condition. Boxed advertisements are available at $15 per inch. Size of type may be specified by advertiser. In answering advertisements all box numbers should be addressed care of Chemical Engineering Progress, Classified Section, 120 East 41st Street, New York 17, N. Y. Telephone ORegon 9-1560. Advertisements for this section should be in the editorial offices the 15th of the month preceding the issue in which it is to appear.

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PEOPLE

(Continued from page 69)

Hiram S. Lukens, Blanchard professor of chemistry at the University of Pennsylvania, retired from active teaching June 30 after forty-six years on the faculty. During these years his various assignments included: national president of the Electrochemical Society, acting dean of the University's Towne Scientific School, director of its department of chemistry and chemical engineering, and head of the chemistry department.

P. Wenzell, Jr., has joined the development department of Celanese Corporation of America. Previously he was with the United States Bureau of Mines, engaged in synthetic fuel development.

Ralph E. Peck, professor of chemical engineering at Illinois Institute of Technology, has been appointed director of the department, effective Sept. 1. He was graduated from the University of Minnesota, receiving his B.S. degree in 1932 and his doctorate in 1936. Between 1936 and 1939 he was an instructor in chemical engineering at Drexel Institute, and then joined the staff of Illinois Tech as instructor of chemical engineering, becoming professor in 1951. Dr. Peck was one of a group of scientists who went to Costa Rica in May, 1951, to do research work designed to boost the production of Manila hemp.

Robert L. Taylor has been appointed director of public relations of Mathieson Chemical Corp. with headquarters at the corporation's executive offices in Baltimore. Mr. Taylor was formerly vice-president of Hill and Knowlton, Inc., New York. He is a past officer of the Manufacturing Chemists' Association, Inc., and for a number of years was editor of Chemical Industries, now Chemical Week. Prior to that he was associated with Monsanto Chemical Co.

Bernard L. Schulman has joined the staff of Esso Laboratories, Standard Oil Development Co. Dr. Schulman completed his undergraduate work at Vanderbilt University, from which he received his bachelor degree in chemical engineering. His doctorate in chemical engineering was granted by the University of Cincinnati.

The appointment of **Harry B. McClure** as executive vice-president of Carbide and Carbon Chemicals Co. was recently announced. Mr. McClure has been vice-president since 1944. He became a member of the organization in 1928 as a research fellow at Mellon Institute of Industrial Research in Pittsburgh. Later he became technical representative in the Philadelphia area and was transferred to sales work in New York in 1931. For the past twenty years he has been concerned mainly with the development of new chemicals and finding new industrial uses for them. In 1950 Mr. McClure received the first honor award of the Commercial Chemical Development Association for valuable service to the chemical industry.

John S. Zinsser, vice-chairman of the board of Sharp and Dohme division, Merck & Co., Inc., Philadelphia, has been elected to the Committee for Economic Development, a research and educational organization of business leaders and educators.

Martin de Simo has joined Chemical Construction Corp. as assistant sales manager. He had been technical director of sales of Foster Wheeler Corp. since 1950. Earlier he was an independent consultant in Chicago for two years and vice-president in charge of research at Great Lakes Carbon Corp. from 1941 to 1948. He was associated with Shell Development Co. from 1928 to 1941, first as a research chemist and later as head of the organic chemicals and petroleum processing departments. Dr. De Simo was graduated from the Technical University at Zurich, Switzerland, in 1926. He received his doctorate in chemistry from the University of Lausanne in 1927 and pursued further chemical studies at University College, London, England, from 1927 to 1928.

Robert F. Dye is now associated with the research and development department, Phillips Petroleum Co., Bartlesville, Okla. He received his Ph.D. from Georgia Institute of Technology.

The recent appointment of **Wilfred Gains** as assistant to the vice-president in charge of engineering, Catalytic Construction Co., has been announced. Mr. Gains was graduated from McMaster University, Hamilton, Ontario, with a degree in chemical engineering. For twelve years he was associated with the Chemical Construction Corp. as a senior chemical engineer in charge of the design, construction, and operation of am-

monia and other petrochemical plants. More recently he was a senior project engineer with the Vulcan Copper and Supply Co.

Donald C. Lewis, formerly in the research and development department of the Merrimac division, Monsanto Chemical Co., is now in the sales department of that division. He has been assigned to the industrial chemicals department as technical service representative, with major interest in chemicals for the paper industry. Mr. Lewis graduated from Worcester Polytechnic Institute with a B.S. degree in chemical engineering in 1951.

Max F. Mueller, formerly chief engineer for the J. T. Baker Co., has been named chief technologist of Grace Chemical Co.'s nineteen-million-dollar nitrogen plant now under construction near Memphis. Temporarily he will be assigned to the development department until he takes up his new duties. Since graduation from the University of Michigan as a chemical engineer, he has been engaged in project and process engineering in the chemical industry in Michigan, Maryland, and Pennsylvania.

The retirement of **Francis L. Mead,** who was responsible for many improvements in the current processes used for production of citric acid by Chas. Pfizer & Co., was announced recently. Mr. Mead joined Pfizer in 1920 during the period when the firm was pioneering in the production of citric acid by the fermentation process. During his long career with the company, he served as manager of citric acid recovery and more recently as new products manager. Mr. Mead was in considerable demand as a consultant and was loaned to other firms in Europe and Canada to aid in the construction of citric acid facilities there.

The appointment of **Russell L. Miller** as production manager of Monsanto Chemical Co.'s Merrimac division was recently announced. Mr. Miller was first employed by Monsanto in 1936 as a chemist in the William G. Krummrich plant at Monsanto, Ill. In August, 1947, he left the Krummrich plant, where he was serving as operating superintendent, to become plant superintendent at Everett. He was made Everett plant manager in March, 1948. Mr. Miller holds a B.S. degree in chemical engineering from Iowa State College.

Necrology

C. C. AHLUM

C. C. Ahlum, long associated with the Du Pont Co., died recently. He was graduated in chemistry from the Drexel Institute, Philadelphia, and early in his career was employed by George W. Lord Co. of Philadelphia as analyst.

INDEX OF ADVERTISERS

Advertising Offices

THE PRESIDENT SAYS

. . . A.I.Ch.E. meetings at scattered points are a good thing . . .

Until comparatively recent times National meetings of A.I.Ch.E. were held only twice a year. A meeting in the spring was known as the semi-annual meeting and the fall meeting was the annual meeting, stipulated by the Constitution as the annual business meeting.

The growing size of our membership and particularly the rapid increase in the number of young members inspired a plan for regional meetings. The idea was to arrange each year at scattered points a number of meetings which people could attend without having to travel long distances. Keeping these regional meetings comparatively simple would reduce costs to the point where large numbers of our members could afford the transportation and fees. Since business organizations and educational institutions finance only just so many convention trips, we expected to serve through individually financed regional meetings many more folks than would otherwise be possible.

The resounding success of the early regional meetings really defeated the original purpose of the plan, in my opinion. These meetings were so well attended and drew from so wide an area that they never were truly *regional* in character. *Regional* became a misnomer, as we simply switched to *four* national meetings instead of two. The costs involved were neither as originally intended, nor compatible with the limited financial resources of young men. I have always regretted that we did not detect the trend of events much sooner and do something to recapture the original spirit of the plan. I am certain that the basic principles were correct and that the need which existed then exists now, very likely to a greater extent.

I consider it necessary that chemical engineers be given an opportunity to foregather occasionally to listen in a group to papers on chemical engineering and to discuss them in session and out. I believe it is a valuable experience for a young chemical engineer to have a chance, occasionally, to rub elbows with some of the great men of the profession, men with whom he is not likely to come in contact otherwise. I believe that out of such contacts an enthusiasm is born for his field which cannot be gained in any other way, and this sort of thing can be arranged better by A.I.Ch.E. than any other agency.

Many of our young members regularly attend Institute section meetings in their own localities. This is good. Others have no opportunity to do so because they are not near a local section. But in any event, I have in mind something quite different from the ordinary Institute section program. For one thing, I think it makes a great deal of difference whether one is attending an evening session after a hard day's work or attending a meeting lasting one or two days. The entire approach is different. Furthermore, there is virtue in gathering chemical engineers from a wider area than is served by any one section. In several parts of the country, meetings are being held which embody many of the elements of regional meetings, as I think of them. These are arranged by one Institute section in some cases and by several cooperating sections in others. They last, usually, one day. I have attended some of these and they are extremely useful. Unlike the National meetings, all arrangements are made locally without recourse to the Institute's Program Committee or Secretary's office. Programs and invitation are not distributed to all Institute members. Attendance is almost entirely from surrounding territory, although sometimes there are visitors from afar who have somehow learned about a program which holds special and possibly pressing interest for them. If many people from remote spots began to attend these meetings, then the entire complicated setup for handling hotel accommodations, entertainment, and transportation problems would become necessary and complexity would lead to expense, programs would become more elaborate in order to justify long trips, and the local man of limited financial resources would be frozen out again. It seems to me an inescapable fact that complex meetings are not necessary to the principal purposes of getting together and that complexity is especially undesirable in what I think of as regional meetings.

I should like to see a great many more truly regional meetings. Arranging one is not a tremendous task and is easily within the abilities of a single Institute section. If two or more sections are close enough together to cooperate effectively, so much the better. This is being done by sections in Ohio and western Pennsylvania, for example, and in other areas. In my opinion such meetings can serve a useful purpose if they *deliberately avoid trying to imitate national meetings*. Rather, they should be geared to the purses and the time schedules of members of our profession who cannot spare large amounts of either money or time away from the job. It has been my observation that almost all employers are glad to allow time to their workers to attend really worth-while sessions if only a day or so is involved, whereas only a small percentage of chemical engineers can be spared for a longer period. I have also noted that many employers will even underwrite expenses or partial expenses for comparatively large groups of chemical engineers, provided the cost per head is relatively insignificant. Chemical engineers are willing to finance themselves if necessary and if they believe the meeting is worthwhile.

I urge every Institute section not now active in this field (and every one not active enough) to consider seriously undertaking alone or in cooperation with other near-by sections, one or more regional meetings each year. I am certain that this would be of great service to our members and would bring into our fold a good many men who ought to be a part of A.I.Ch.E.
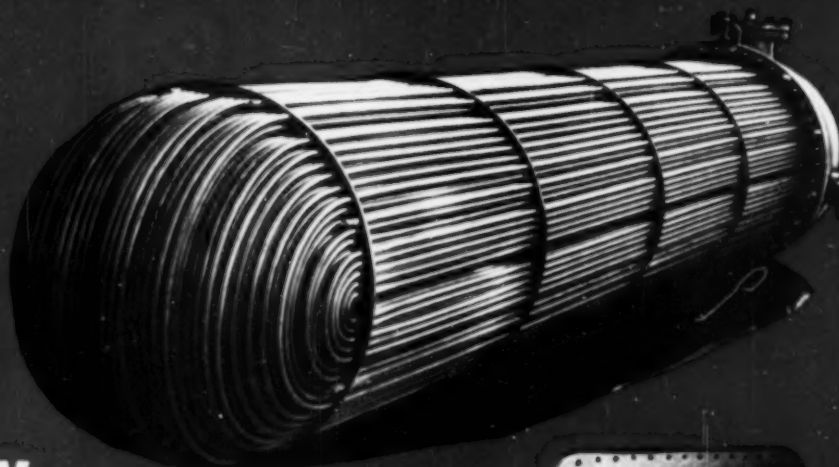
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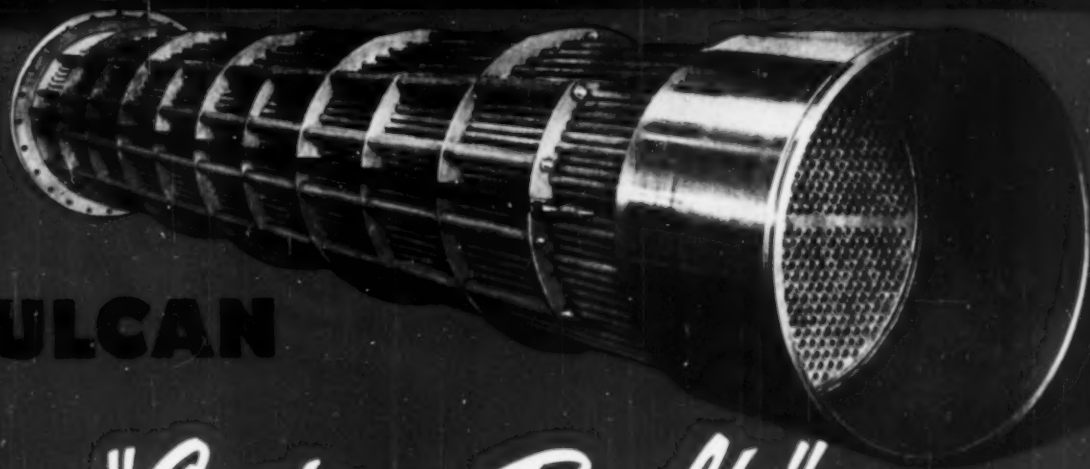
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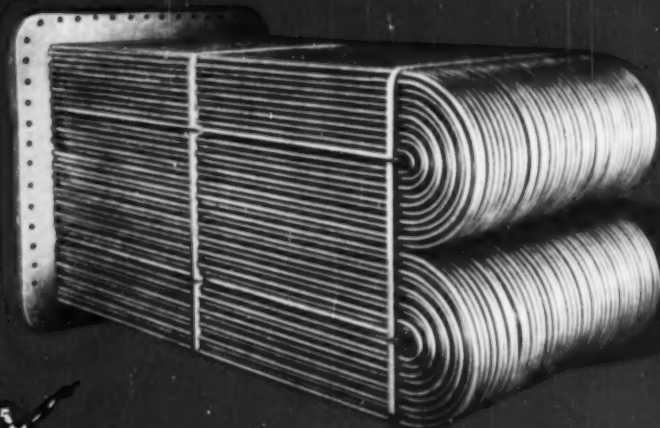
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